A SYSTEMATIC APPROACH TO CARBON FOOTPRINT REDUCTION STRATEGIES IN INDUSTRIAL PARKS

A Thesis

by

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ABSTRACT

Global concerns of the effects of increased carbon dioxide emissions have pushed policy makers to adopt strict emission targets .The introduction of such targets will require strategies to achieve them in efficient ways. In Qatar, the world highest per capita carbon emitter, stationary emission sources are responsible for the majority of the country's footprint. Concentrated in industrial parks or clusters emissions cuts would be more significant and efficient to achieve as part of a national emission reduction strategy. Conventional mitigation techniques involve fossil fuel replacement, increasing energy efficiency and the use of carbon capture and sequestration. This research proposes an alternative method based on utilization or conversion of carbon dioxide within the industrial city. Carbon dioxide can be chemically or biologically converted into fuel, polymers, food supplements, fodder or another value added product could be used in applications of enhanced oil and gas recovery. Based on concepts of classical network design and Industrial Ecology (IE), a systematic approach was developed to obtain a cost-optimal network of carbon reduction. The integrated analysis of utilization options together with the capture, separation, compression and transmission of carbon dioxide will be required to determine the most economically attractive footprint reduction solutions. The optimization was illustrated using a case study.

DEDICATION

To my mother and my grandmother.

ACKNOWLEDGEMENTS

I am sincerely in debt to the various people who have helped me throughout the course of my studies. I owe a great deal of thanks to the following individuals, who have been wonderful mentors, contributors and supporters:

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NOMENCLATURE

ε_k	Sink efficiency factor
ε_T	Treatment unit carbon removal efficiency
Fs	Total raw source available flow
Fsi	Flow of treated source
Fsj	Flow of untreated source
ysi	Weight composition of the treated stream
ys	Weight composition of the raw source
L_s	Minimum total flow out of raw source
U_s	Maximum total flow out of the raw source
$F_{si,k}$	Flow from treated source Is to sink k
$F_{sj,k}$	Flow from untreated sour Js to sink k
F_k	Total flow into sink k
y_k	Weight composition of sink k
$y_{s,j}$	Weight composition of untreated source Js
G_k^{max}	Maximum flow into sink k
Z_k^{min}	Minimum concentration required at sink k
$L_{si,k}$	Lower flow limit of the treated source Is to sink k allowed in pipes
$L_{sj,k}$	Lower flow limit of the untreated source Js to sink k allowed in pipes
$U_{si,k}$	Upper flow limit of the treated source Is to sink k allowed in pipes

$U_{sj,k}$	Upper flow limit of the untreated source Js to sink k allowed in pipes
$X_{s,ik}$	Binary associated with flow of the treated source (Is)
$X_{sj,k}$	Binary associated with flow of the untreated source (Js)
$C^{Pipe}{}_{si,k}$	Pipe cost per unit length of the connection between treated (Is) source to matched sink (k)
$C^{Pipe}{}_{sj,k}$	Pipe cost per unit length of the connection between untreated (Js) source to matched sink (k)
$C^{T}{}_{si,k}$	cost of treatment applied on the treated source treatment cost as in carbon removal
$C^{sink}{}_{k}$	Cost of the sink (k) processing used
$C^{comp}{}_{si,k}$	Total annualized cost of compression source Is to meet sink (k) requirements
$C^{comp}{}_{sj,k}$	Total annualized cost of compression source Js to meet sink (k) requirements
$C^{pump}{}_{si,k}$	Total annualized cost of pumping source Is to meet sink (k) requirements,
$C^{pump}{}_{sj,k}$	Total annualized cost of pumping source Js to meet sink (k) requirements
$H_{s,k}$	Distance between raw source (s) and sink (k)
CRF	Capital Recovery Factor
n	Number of years, equipment life

i	Discount rate
$CC^{capital}$	Compressor Capital Cost
$CC^{Operating}$	Compressor Operating Cost
$\Delta P_{si,k}$	Pressure difference between source Is and sink k
$\Delta P_{sj,k}$	Pressure difference between source Js and sink k
$\Delta P^{pipe}_{si,k}$	Pressure drop between source Is and sink k
$\Delta P^{pipe}_{sj,k}$	Pressure drop between source Js and sink k
η	Efficiency of the unit
ρ	Density of the stream
W_{si}^{p}	Power of pump of source Is
W_{sj}^{p}	Power of pump of source Js
$D^{c}_{si,k}$	Commercial diameter of pipe between source Is to sink k
$D^{c}_{sj,k}$	Commercial diameter of pipe between source Js to sink k
$D_{si,k}$	Calculated diameter of pipe between source Is to sink k
$D_{sj,k}$	Calculated diameter of pipe between source Js to sink k
T_{si}	Temperature of stream outlet at source Is
T_{sj}	Temperature of stream outlet at source Js
M_{si}	Average molecular weight of stream from source Is to sink k
M_{sj}	Average molecular weight of stream from source Js to sink k
v	Average velocity at outlet of the source
v_{si}	Average velocity at outlet of the source Is
v_{sj}	Average velocity at outlet of the source Js

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1. INTRODUCTION

The consensus amongst climate research scientists attributes climate change to anthropogenic activity. In 2010, annual global Greenhouse gas (GHG) emission have reached 49.5 billion tons of CO_2-equivilant, a level higher than any previous year (Edenhofer et al, 2014[1]). This trend is expected to continue with the increase in global population [2]. The United Nations projects that by the year 2050, the earth's population would reach 9 billion people, which will subsequently, increase the demand for energy, food, water and fresh resources (Metz et al [3]). Policy makers responded by introducing limits on carbon dioxide (CO_2) as it is the largest component of anthropogenic GHG emissions. This creates challenges for the energy intensive industrial sectors to manage their carbon footprint.

Carbon dioxide can occur naturally in nature, but the majority of the CO_2 released is due to human use of fossil fuel in power and heat production, transportation fuel consumption, or released as a by-product of industrial activities or agriculture. Globally, stationary sources are responsible for about 60% of carbon release of the total fossil fuel footprint ($23GtCO_2$/yr in 2000) [3]. This has led to much global and regional effort aiming to reduce carbon footprint. The Intergovernmental Panel on Climate Change (IPCC) recommended a target of 30% for developing countries and 80% for industrial countries to be reached by 2050 in order to avoid dangerous climate change effects [3]. Although defining dangerous climate change effects can be open for interpretation, scientists agree it would be more severe than the hurricanes, floods and

other weather changes that have taken place (Metz et al, 2007[4]). The European Union set 20% target by 2020, to be increased to 80% by 2050 (EU Climate Action [5]). The UK similarly adopted a proposal to reduce its carbon footprint to 80% by the year 2050 (Department of Energy and Climate Change [6]). Even though Qatar has not adopted a similar doctrine yet, it did take positive steps towards showing environmental commitments. Globally Qatar is ranked the highest per capita producer of carbon dioxide emitting an average of 40.31 metric tons per capita (World Bank [7]). Through Qatar National Development Strategy (QNDS), it established a program that supports growth under environmental management (QNDS [8]). Since its drafting in 2011, Qatar was able to produce a policy that manages air pollution and GHG [9]. Moreover, Qatar played a role in global emission reduction efforts by hosting the United Nations Framework Convention on Climate Change (UNFCCC) meeting Conference of Parties COP 18 in 2012. Therefore, the need for CO_2 mitigation methods is of importance in the region. Since the main sources of CO_2 emissions in Qatar are related to the industrial sector, it is clear that the stationary sources are responsible for more than 70% of the total country's emissions as shown in Figure 1.

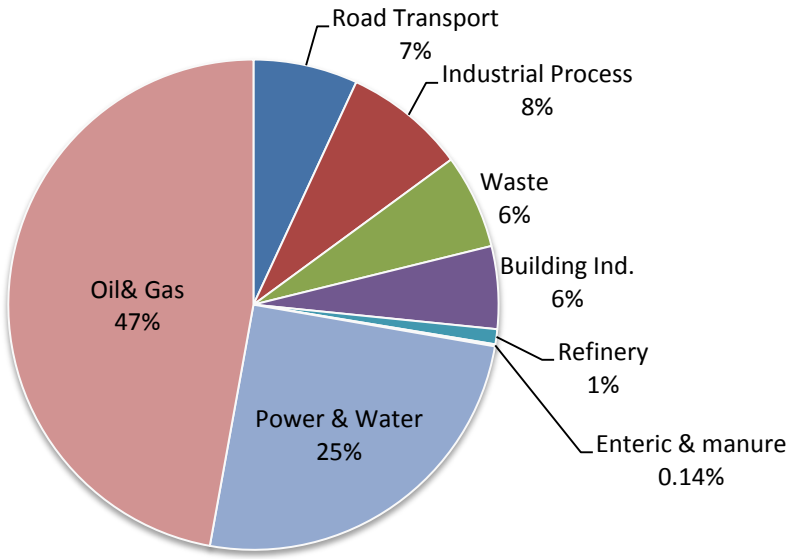

Figure 1: Qatar Main Emission Sectors (Global Institute [10])

Carbon dioxide can be reduced in one of the three methods, reducing energy and carbon produced from combustion by energy integration, reducing carbon from the fuel itself by the use of low footprint fuels and carbon storage and sequestration. Whilst the use of clean and renewable energy sources, energy efficiency and carbon capture and storage (CCS) has received much attention from the scientific community, little work has been disseminated to date on approaches to explore carbon utilization options. Carbon dioxide can be utilized in many different ways, including chemical or biological conversion into fuel, polymers, food supplements, fodder or another value added product. In addition, storing carbon dioxide underground might also create economic incentives to cut emissions in case of Enhanced Oil Recovery (EOR). The integrated analysis of utilization options together with the capture, separation, compression and

transmission of carbon dioxide will be required to determine the most economically attractive footprint reduction solutions.

This work presents a systematic approach to integrate carbon dioxide within an industrial park with multiple carbon emitting streams and the potential carbon utilization options (sinks) that may exist. The approach identifies the lowest cost integration options to attain a given footprint reduction target and decides which sources are best captured together with their allocation to the various utilization options. It determines if the captured carbon dioxide should be purified and takes into account costs of carbon capture, carbon transportation between sources and sinks as well as costs associated with the utilization processes.

A literature survey is presented in section two of the most relevant carbon mitigation strategies. Section three focuses on the addressed problem and provides background to the developed approach. The overall approach is subsequently described in section four followed by the optimization problem statement in chapter five. Section six presents the mathematical model implemented while section seven applies the developed strategy and model on a case study. Finally, section eight presents the overall conclusion and recommendations for future work.

2. LITERATURE REVIEW

This section summarizes insights into carbon reduction methods that apply concepts of energy and mass integration for carbon reduction across plants. It also looks at pervious work in terms of the carbon allocation problem in addition to exploring eco-industrial parks design approaches that applies the concept of re-use and waste elimination in a systematic method.

2.1 Mass and energy integration for carbon reduction

Energy integration on an individual process or across multiple units has been applied to reduce carbon footprint. Many strategies used graphical methods to apply energy integration in order to achieve carbon emission target. Within a single plant graphical techniques have been applied first by Smith and Delaby [11] by relating emissions with energy targets. They focused on reducing the amount of emissions generated from utility systems and compared footprints of gas turbines versus steam boilers [11].More work has emerged in terms of solving the problem of carbon dioxide generation from an energy integration point of view that later expanded to include multiple processes. Klemes et al [12] applied a graphical method on a total site utility system to save fuel thus cutting global emissions. The method achieved more cuts globally rather than looking at individual processes. Emission reduction by energy integration across multiple plants was also addressed graphically by Jia et al [13]. They applied pinch analysis on an industrial park to determine the energy target and the

corresponding emission. These methods would aim to produce optimal energy use with minimal emission, but that might not be the best case economically .To overcome that Al-Mayyahi et al [14] incorporated economics using a bi-objective optimization of utility generation and carbon emissions. The method graphically explored the trade-offs between costs and emissions to reach an optimal solution [14].

Fuel replacement and fuel switching from fossil fuels to less carbon intense fuels or renewables have been explored as part of the general integration problem. Graphically, Tan and Foo [15] applied a Carbon Emission Pinch Analysis (CEPA) targeting method that minimized emissions from different energy resources. It was developed to identify and allocate the minimum number of zero-emission energy sources required to meet the carbon target while adhering to the energy demand [15]. Renewables, nuclear energy and biomass as well as fossil fuels with carbon capture application were defined as the zero-emission energy sources to replace more carbon intense fuels. Using a total site targeting methodology coupled with fuel replacement was incorporated into the energy integration mix by Perry et al [16]. They applied renewable energy e.g. solar cells, biomass with typical engines and heat pumps, at the same time as considering residential and service sectors energy consumers as part of the total site target [16].

Graphical methods have been applied as well with the use of carbon capture and storage (CCS) as means to reduce carbon footprint associated with energy use. This is illustrated by the work of Ooi et al [17] that falls under the umbrella of the emerging Carbon-Constrained Energy Planning (CCEP) problem. While graphical methods give

good insights to the carbon reduction and energy planning problem, mathematical approaches can overcome simplifications present in graphical approaches that might limit its wider applicability. Combining energy integration and fossil fuel replacement Hashim et al [18] addressed the problem of emission reduction from power generation grid using a Mixed Integer Linear Program (MILP). Source-sink representation to deal with energy integration and carbon footprint targeting was applied by Pekala et al [19]. Their method used carbon capture and storage as interceptor as in mass integration problems to purify energy sources as well as bio-fuels as carbon-neutral sources. Likewise, Shenoy and Shenoy [20] took carbon capture into consideration and applied energy integration with carbon removal. This approach looked at overcoming the energy penalty associated with carbon capture while incorporating carbon capture to a plant. It applied a hybrid targeting and optimization method using an algorithm for targeting and Linear Programming /MILP mathematical programming for the optimization [20]. On the other hand, combining energy integration with carbon dioxide placement in consuming processes have been addressed by Munir et al [21] through a graphical pinch method coupled with a set of heuristics to reduce emissions based on source-demand system. Reduction included replacing fuel sources with renewals and direct re-use of carbon dioxide placement in demands, using thus a waste-to-resource approach. However the approach did not elaborate on the mechanism behind the individual connection or on the associated cost, both required to determine connection if applied on real systems.

Extensive work has been applied in terms of emission reduction in energy planning or through energy integration. Methods were aimed to reduce carbon dioxide generation as results of combustion thus targeting less energy use and more efficient designs. Other methods looked at cutting carbon dioxide from the source itself while maintaining the energy requirement. Nevertheless, these methods try to limit carbon production; they do not deal with the inevitable produced carbon. While the prospects of applying renewables into the energy grid is appealing, the move in industry towards their adaptation might be slow as fossil fuels are currently the most used energy sources. Therefore, there is a need to apply a method that resolves the carbon produced issue. This has been demonstrated through the study of carbon capture allocation between sources and sequestration sinks.

Attempts to allocate carbon dioxide in utilization sinks across processes have not been studied before. The concept of using the waste or emissions elsewhere as in natural ecological systems stems from Industrial Ecology's (IE) Eco-Industrial Parks (EIP). Designing eco-industrial parks have been the subject of much research literature. Attempts included mathematical models such as inter-plant water integration presented by Lovelady and El-Halwagi [22] and Rubio-Castro's et al [23] program for optimal treatment selection in water based exchanges. Similarly water exchange was studied by Aviso et al [24]. Multiphase water integration in industrial parks was presented by Bishnu et al [25]. While Alnouri et al [26] formulated a model that deals with piping connections in EIP and integrating stream. Interplant energy integration was studied by Chae et al [27], which considered waste heat reduction while Stijepovic and Linke [28]

looked the problem of energy recovery and re-use from multiple plants (Stijepovic et al, 2012[29]).

The use of CO_2 as a synergy method has been previously discussed by Norstebo et al [30]. They studied the case of an industrial city with multiple sources of CO_2 being captured in one unit. Their work was followed by cost study where the amount of flue gas is varied and it showed multi source capture feasibility (Midthun et al [31], Perez-Valdes et al [32]). This was considering CO_2 as material flowing from various points into a capture unit; it did not consider CO_2 allocation. CO_2 placement in processes has been demonstrated before in applications of Eco-Industrial Parks such as Kwinana Australia Beers [33]. However, these attempts were ad-hoc synergies and there has not been an attempt to allocate carbon dioxide systematically between the sources and utilization sinks within a city as far as our knowledge. Work has been demonstrated in matching carbon dioxide sources and sinks in terms of sequestration and enhanced oil recovery as explained in the next section.

2.2 Carbon dioxide allocation

Connections mechanisms and steps involved carbon allocation have been studied previously in terms of enhanced oil recovery and storage sinks. Using a non-linear mathematical model Turk et al [35] were able to represent a network of sources, mainly power plants, and sinks, enhanced oil recovery locations, while accounting for transportation and injection costs. Carbon dioxide delivery and allocation has been studied by Middleton and Bielicki [36]. They used a spatial model mixed integer linear

program (MILP) that dealt with infrastructure of carbon capture and storage. Similarly cost optimal CO_2 was explored by Weihs and Wiley [37]. Knoope et al [38] studied transporting carbon dioxide in different phases. The Multiphase carbon storage was studied by Tan et al 2013 [39], He et al 2013 [40]. Tan et al applied source sink mapping in carbon capture and storage with time, injection and capacity constraint. He et al [40] used followed with a formulation that addressed the robust source-sink matching in carbon capture and storage supply chains under uncertainty. It incorporated elements of physical and temporal constrains in a MILP to discuss CO_2 allocation. Their formulation focused solely on the case of geological storage as sinks and aimed at maximizing CO_2 storage with the possibility of one sources connected to one sink depending on the flow and storage limit.

Enhanced oil recovery sinks matching was studied by Hasan et al [41] on a national scale. They considered a large scale CO_2 supply chain network with the selection of the best capture technology for different CO_2 sources while taking into account geographical locations and pipeline transposition using [41]. Similar concept was applied by Alhajaj et al [42] on a regional scale in the UAE.

Several studies pairing of sources with geological-sinks based on specific site information. Using a scoring strategy, Pasarai et al [43] used a three-step procedure that starts with source ranking based on a 14-criteria set that is followed with a similar ranking of sinks based on capacity and other factors. The final step considers geographical proximity in making connections that link the highest purity sources with the best sinks. Several sources might be used at a time. Similarly, Kobos et al [44]

studied captured carbon dioxide placement and costs in storage sites within the Southwestern United States using data-based software. Yousefi-sahzabi et al [45] explored spatial and environmental feasibility of sources to geological storage connections by applying Geographical Information System (GIS) simulation. It simulated emissions on the source and sinks sites before and after connections and was able to estimate costs. These models show the diversity and specific site characteristics that geological storage entails that cannot be generalized.

These studies were limited based on the consideration of one type of sink process and in terms of connections. Most work connected one source to one sink, avoided mixing of sources at sinks and was constrained by the site specific information. The method proposed in this work addresses the gaps presented in the previous mentioned work in terms of dealing with produced carbon dioxide, type of sinks proposed and the incorporation of spatial and physical elements. It focuses on tackling the carbon reduction allocation in an industrial city using CO_2 converting sinks. The next section presents the focus of the work and illustrates the approach to achieve efficient carbon footprint reduction on an industrial city.

3. SCOPE AND OBJECTIVE

The focus of the research is to develop a systematic method to determine cost effective carbon reduction strategies for total carbon dioxide emission target on an industrial city. Looking at an industrial city emission sources, as shown in the Figure below, emission can arise from either the city's fuel input or due to process inefficacies. As a result, emission reduction can take place at the city input as part of inherent prevention of carbon production, within the city as part of process enhancement to lessen energy consumption or at the outlet of the city as end-of-pipe control measure. The three tactics, illustrated in Figure 2, are explained as follows:

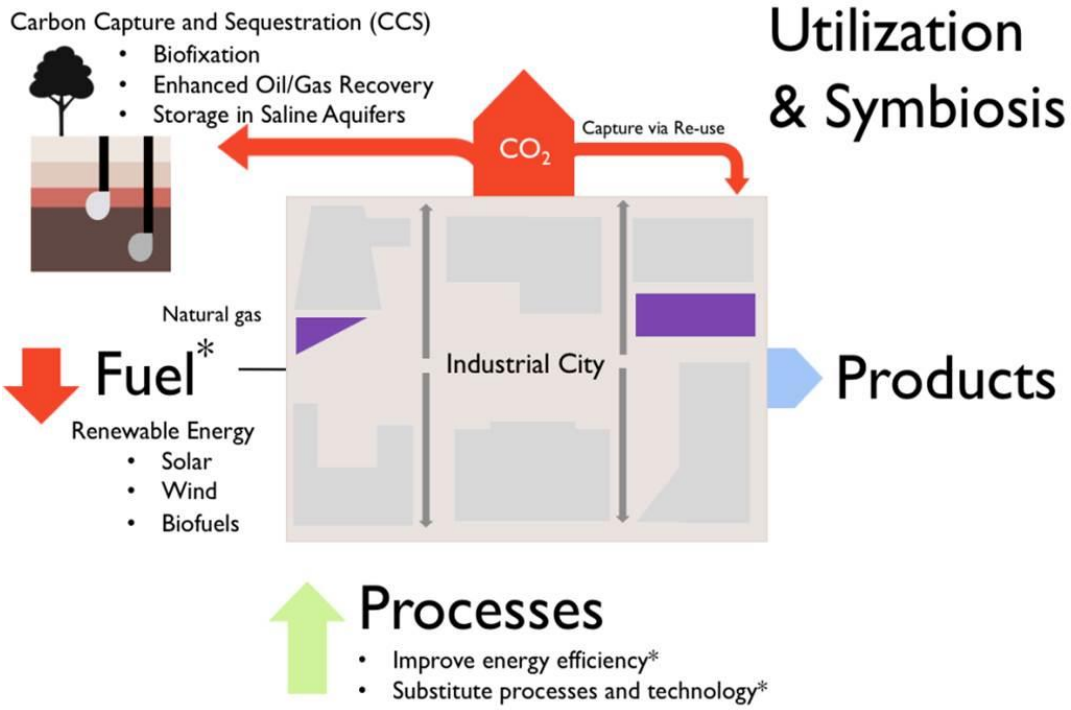

Figure 2: Research Focus, *Not addressed

1. Applying the concept of inherent prevention means that carbon dioxide generation is eliminated. This can be attained by reducing the amount of carbon dioxide generated as a consequence of the fuel combustion via the use of cleaner, less carbon intense fuel or the use renewable energy.
2. Process enhancement within the city is reached by increasing the processes efficiency to generate more of the desired products and less of the unwanted CO_2 by-product or by reducing the amount of energy required for the desired production rate.
3. The last emission reduction method is embodied in carbon capture and sequestration where it is applied as an end of pipe treatment before the carbon stream is discharged or stored. Sequestration can be either biological e.g. natural carbon cycle or geological as in applications of enhanced oil or gas recovery, enhanced coal bed recovery or storage in depleted reservoirs or saline formations.

The elimination of carbon dioxide production in case of power and heat production can be achieved by renewable energy as shown in Figure 2. Renewable energy can be in the form of bio-fuels, solar or wind. Bio-fuels are more effective in case of transportation, while solar and wind powers are used to generate heat and electricity. Increase of technology or process efficiency by the use of energy integration is done to reduce carbon dioxide footprint from the excess usage of fossil fuels. This is applied industrially to save energy resources and reduce the carbon footprint, or to fully use the

carbon feedstock and eliminate carbon dioxide release as by-product. This is limited by chemical reactions and technology. Advances in kinetic engineering and catalysts industry will play a huge role in this area. However, it takes time for designs to be applied commercially in industry.

Currently, the most effective method is carbon dioxide geological sequestration as mentioned earlier, shown in Figure 2. Limitations exist in terms of site information and storage locations that might hinder the adaptation of CCS on industrial cities. Moreover, the costs associated with carbon removal technology are still significant.

The research focuses on an alternative method where carbon dioxide can be contained within the city itself by recycling the carbon emitted back to city as shown in Figure 2. Through applying the concept of utilization, carbon dioxide would be treated as a resource rather than a waste. Different streams of carbon dioxide would be present that can be purified using carbon removal technology. Type of products and processes required would need to be generated to best suit the set-up of the city. The city itself might contain existing processes that can consume carbon dioxide. Further, the logistics of making synergies within the emission points and the consuming processes would need to be determined as explained in the next chapter.

Carbon dioxide has many applications in industry. It can be converted chemically to various fine chemicals, liquid fuels and polymers. It also can go through biological conversion as mentioned earlier via photosynthesis in terrestrial plants to give a number of products from greenhouse produce such as cucumber and tomatoes, or to algae extracted omega-3 fatty acids and proteins. It can go through natural chemical reactions

such as mineralization or physical transformations to fit industrial applications such as supercritical solvents and welding. Mikkelsen et al [46] offers a review of many conversion routes. The decision of conversion route is open to many variables. This imposes several questions that need to be addressed.

- Will the industrial city be able to consume carbon dioxide, convert or store CO_2? If not, what plants or processes should be added?
- Does the process needed fit within the city structure? What are the aspects of a good process? A good emission source? A good sink?
- What technology can be used to consume carbon dioxide? What will be the desired product? What are the capacities needed? What are the requirements needed to be processed through a sink?
- Which carbon sources can be used directly for re-use within the city? If not, how can they qualify to be used? What are the parameters required to determine which sources to capture and the needed processing sinks?
- How much carbon flow is needed to achieve a certain reduction target?
- How does CO_2 conversion look in terms of other methods such as geological storage, utilization through oil and gas recovery?
- How will the connections be made? Which sources to mix in order to achieve sinks requirement?
- How does the cost profile look like? What are the major cost elements needed to evaluate the effectiveness of the process?
- How would carbon delivery take place? Will location play a role?

Therefore, the aim is to develop a strategy that would be able to accommodate those concerns. The strategy would aim to be applied to retrofit or grass-root industrial cities aiming at meeting a certain carbon reduction target. It can be applied on a number of sources and whole cities. It can be applied on a national scale in conjunction with policy makers to reduce footprints. The next section presents an overall approach that attempts at answering the addressed questions.

4. OVERALL APPROACH

The reduction method is designed to take place in an industrial city, which contains plants and processes. Illustrated below in Figure 3 is a representation of the city. In order to describe the approach, the terms process, plant and industrial city are defined below as adopted from Stijepovic and Linke [28].

- A *process* is defined as sequence of operations that convert raw materials into products and produces CO_2 as a result of consuming utilities or as by-product.
- A *plant* is defined as an independent production unit which consists of one or more processes.
- An *industrial city* is characterized by a set of plants located in a centrally administered area of concentrated industrial activity.

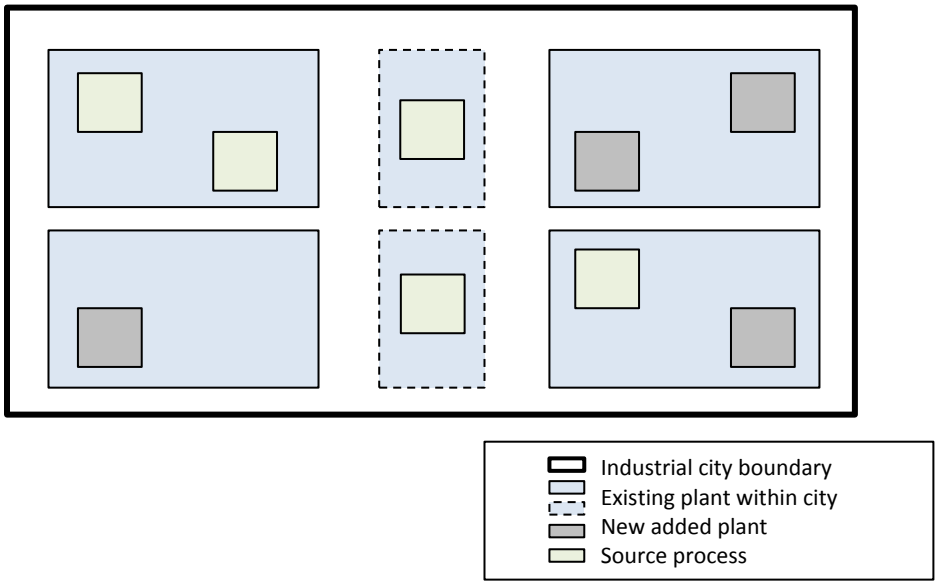

Figure 3: Schematic Representation of an Industrial City Layout

Sinks can be a process within the industrial city or a sequestration application such as biological or geological sequestration. Sinks are defined as units that intake, consume, convert, transform or store carbon dioxide. Carbon dioxide used in enhanced oil/gas or coal bed recovery is also considered a geological utilization application. A source is defined as a stationary emission point of a process within a plant in an industrial city. Source can have different compositions, volumes and parameters. They can be either a process emitted or purged sub-stream e.g. pre-capture reformer unit or an exhaust of a fuel combustion in a furnace or a boiler. Table 1 summarizes the typical emissions that can be found in an industrial city as adopted from IPCC [3].

Table 1: Typical Carbon Dioxide Concentrations in Industrial Applications

Source	CO_2 Concentration vol% (dry)
Power station flue gas :	
Natural gas fired boilers	7-10
Gas turbines	3-4
Oil fired boilers	11-13
Blast gas furnace	
Before combustion	20
After combustion	27
Cement kiln	14-33
Integrated gasification combined cycle synthesis gas	8-10
Ammonia production	18
Ethylene Oxide	8
Hydrogen production	15-20
Methanol production	10
Natural gas processing	2-65

Even though, plants are contained within the city boundaries and are connected to each other using corridors, sources and sinks can be located on different plants. Similarly, source and sinks processes can be in the same plant, yet they are located at a certain distance from each other. Thus, the consideration of spatial arrangement of the city plays a significant role in the determining the location and size of new CO_2 sinks. This opens ways for new processes such as algae productions or greenhouses. These are land specific and thus would determine the CO_2 capacity intake. More importantly, distances between emission point, sources, and CO_2 consuming processes ,sinks, are important in determining the phase and cost of transporting CO_2.

Carbon integration is the recovery of carbon dioxide sources into sinks in order to minimize the overall carbon footprint of the system. The application of carbon integration on the studied system to attain minimal cost network entails several aspects, mainly separation and treatment, transportation and sinks demands. As shown in Figure 5, the sink requires a certain grade and purity of carbon dioxide in order to function. This is achieved through carbon removal unit that would be associated with a cost. Moreover, sinks requires a given pressure that in turn determines the phase of carbon dioxide delivered e.g. for the use of carbon dioxide in storage application, carbon would need to be compressed to a minimum of 15MPa in order to fit pipeline transportation requirements Knoope [44]. This means CO_2 would undergo a phase change and up to certain point a pump would be needed to transport the liquid carbon dioxide, thus adding an extra cost. Similarly, carbon delivery itself would be associated with a certain pressure drop that would be overcome using compression. All of these elements are

needed in order to properly estimate the cost of a source-sink connection in aim of total carbon reduction.

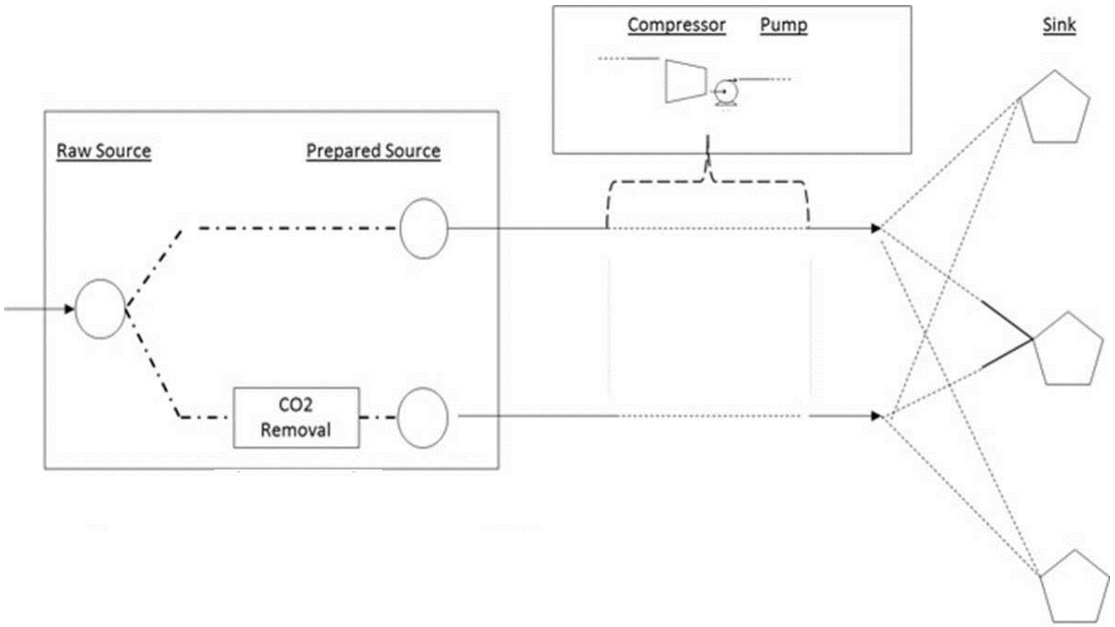

Figure 4: Steps Required in a Connection

A 4-step proposed systematic methodology to reduce carbon footprint on an industrial city by developing economically networks connecting carbon dioxide sources and sinks. The overall methodology is as follows:

Step 1: Data acquisition for industrial zone

Step 2: Identification of carbon sinks

Step 3: Treatment, transmission and cost data

Step 4: Identification and design of cost effective carbon integration network

The first three steps detail information and data that should be acquired in advance in order to obtain a representation of the system, while the last step defines the optimization problem that returns the solution of the best carbon reduction networks.

4.1 Step 1: Data acquisition for industrial zone

In this step relevant data of the studied industrial city are gathered. It requires the knowledge of the following information.

A. Spatial information, in order to estimate the transportation cost would require:

- Plots of existing plants and plots of new plants/sinks
- Corridors

B. Source-Plant data which include:

- Total emission of plants
- Sources of significant flow
- Composition of source stream

4.2 Step 2: Identification of carbon sinks

The purpose of this step is to identify the sink options that can reduce carbon dioxide. This step is dependent on the system developer and the studied industrial city. It entails an extensive literature review for CO_2 utilization technologies. Based on selection criteria, the inapplicable options are eliminated.

Technical data collected include:

- Potential amount CO_2 fixed
- Current application in industry and stage of development
- Location and size of the facility needed
- Approximate cost and possible revenue
- Maximum and minimum flow of CO_2 to sustain operations (scale and capacity)
- Maximum and minimum composition of CO_2 required in addition to impurities tolerance
- Sink efficiency factor ε_k

The sink efficiency factor, ε_k is added to assess any release carbon dioxide from the sinks and is defined as follows

$$\varepsilon_k = \frac{Carbon\ dioxide\ emitted\ or\ released\ from\ sink}{Carbon\ dioxide\ inlet\ to\ the\ sink} \quad (1)$$

The parameter was added to give a realistic insight into the effectiveness and expected duration of carbon dioxide storage of CO_2 in sinks. This is done in order to fully compare different carbon capture options. It can be obtained through literature or estimated based on the sinks fuel requirements and carbon dioxide released sub-streams.

This is followed by a screening process to eliminate inapplicable options to the existing system and reduces the solution space. This is done to reduce the solution space and refine the potential sink options. The selection is based on:

- Process maturity

- Scale of fixation of carbon dioxide
- Potential revenue and market demand
- Homogeneity with the existing plants
- Availability of geographical storage

Maturity considers the technology stage of development and commercial application. Technologies that are still in research phase and have no real potential applications are eliminated. Scale of fixation determines the technologies with the most CO_2 fixation in regard to inlet and conversions. It eliminates solutions that fix the least amount of CO_2. Potential revenue and market demand eliminate the solutions that generate the least income in regard to the other options based on user preference. Homogeneity with the existing processes considers the compatibility of the selected sink with the existing plants in the industrial city in terms of raw materials requirements and type of industry. Availability of geographical storage looks into the sequestration possibilities in the region that would be able to intake the city produced carbon dioxide.

4.3 Step 3: Capture, transmission and cost data

Depending on the source conditions and the sink requirement, the connection involves dehydration, compression and pumping. Figure 4 shows the components involved in the connection. Therefore, the following information is required.

- Treatment units' costs are obtained depending on the chosen technology and available streams chemical/physical absorption, membrane technology or adsorption.
- Treatment units' carbon removal efficiency ε_T which can be calculated based on the carbon recovery of the unit and the heat requirement associated emissions as explained in Appendix A.
- Carbon dioxide transmission costs, which can be estimated based on models, which are widely available in literature. Costs are based on flow of carbon dioxide and can be adjusted according to the pipe size chosen.
- Compression degree and associated costs depend on the source sink pressure difference and the pressure drop in the chosen pipeline. Depending on the pressure costs can be of blowers, turbine, compressor and pumps or a combination.
- Literature survey of common carbon dioxide prices and costs within the market, this is going to act as the capture cost within a sink.

4.4 Step 4: Identification and design of cost effective carbon integration networks

This step is achieved through developing and solving an optimization. The formal problem statement required to solve the model is formulated in the next section.

5. PROBLEM STATEMENT

The general problem addressed in this work is CO_2 placement in a process that converts/stores CO_2 as means to reduce emissions in an industrial city. The main objective is to determine the most economical network for a set of capture targets. The problem is formally stated as:

Given

- An industrial city with a number of plants
- A number of emission streams (sources) of a known flow rate, composition, temperature, pressure and location in each plant
- A number of existing/potential processes that can take in emissions (sinks) with a known CO_2 flow rate capacity, minimum composition, temperature, pressure and location
- Distances between all sources and all sinks

Determine

- The amount of CO_2 flow required between sources and sinks
- The amount of CO_2 purified through treatment, compressed and pumped
- The cost-optimal network the connects sources and sinks within the industrial city for a given carbon footprint reduction

A number of sets are defined as basis for our problem formulation:

S $\{s|s=1,2,3,\ldots,N_{sources}|$ S is a set of carbon sources $\}$

$I_s\{i|i=1,2,3,\ldots,N_{source}|$ I_s is a set of treated carbon streams from source s $\}$

J_s {j|j=1,2,3,…,N_{source}| J_s is a set of untreated carbon streams from source s }

K {k|k=1,2,3,…,N_{sinks}| K is a set of carbon sinks}

The overall source sink mapping of the problem is presented in Figure 5. Sources with a given flow are called raw sources. The amount of raw sources flow (Fs) is determined through the optimization. The raw source is allocated through its spitted two streams, treated (Fsi) and untreated (Fsj). The treated mainly goes through a treatment unit, defined here as carbon removal and emerges a concentrated stream of carbon dioxide with composition (ysi). The untreated sources are a split stream with the same composition as the raw source (ys). Therefore, streams from all sources were assumed to either be treated (in this case, belonging to the subset, Is) or untreated (in this case, belonging to the subset, Js). This new representation based on the classical network design problem as illustrated in Figure 5 is carried to reduce the level of complexity in regard to the treatment. Hence, the computation time is cut short while keeping the effect of the treatment in place.

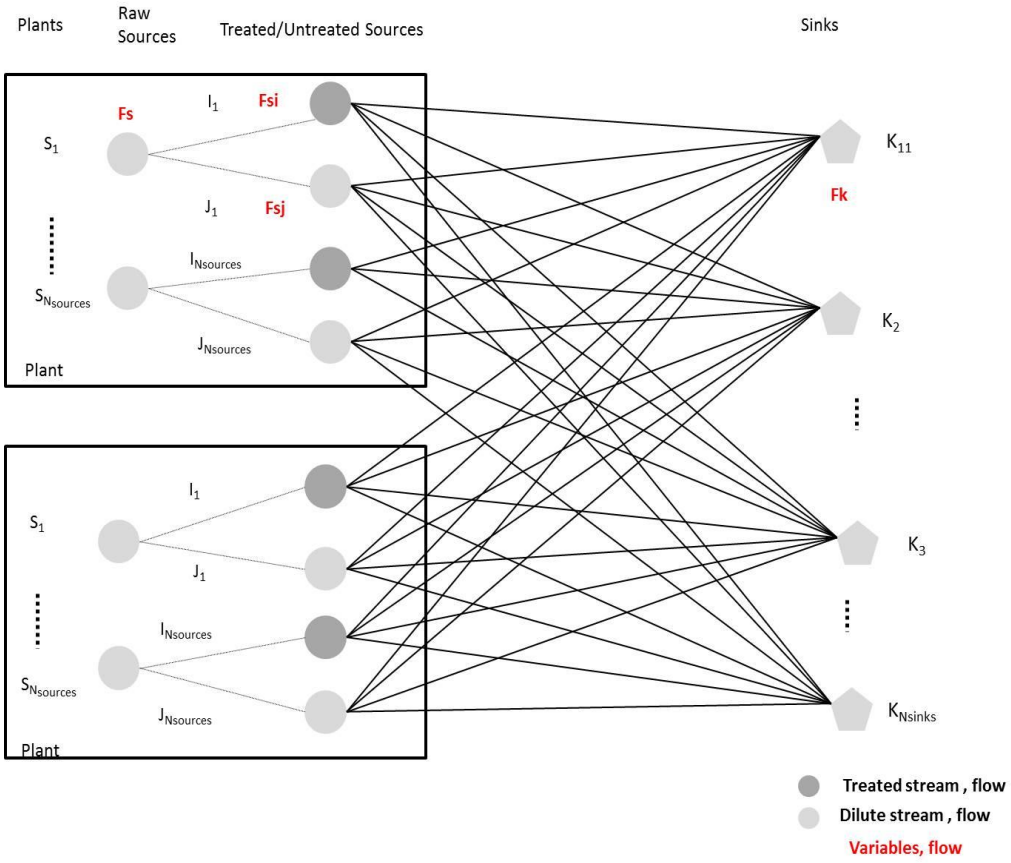

Figure 5: Source-Sink Matching

6. PROBLEM FORMULATION

The optimization formulation developed for minimizing the total cost by carbon dioxide allocation in utilization sinks is presented in this section. The superstructure of the problem is illustrated in Figure 5. The objective is to minimize the total cost of the network including the cost of processing carbon dioxide in sinks, the cost of treatment and costs of compression and delivery. The objective function is described below:

MIN TC (2)

Where, TC is the overall total cost calculated as follows

TC of treated and untreated sources = Cost of sinks + Cost of treatment + Cost of Compression + Cost of delivery

The cost of sinks refers to the price the sinks are expected to pay for carbon dioxide as delivered. Here, a positive number is defined as a cost while a negative number indicates a profit. Cost of treatment refers to the cost of carbon dioxide removal using a traditional carbon capture unit. Compression cost accounts for the total compression and pumping needed to meet sinks requirements and overcome pressure drop due to piping. The cost of delivery is the total piping cost required.

The optimization determines the amount flow required out of the raw source based on an upper and lower limit as follows:

$L_s \leq F_s \leq U_s$ $\forall\ s \epsilon S\ k \epsilon K$ (3)

Here, U_s is the maximum flow available from the raw source and L_s can be set by user based on the system requirements. A total flow balance around raw sources s, ensures that the flow does not exceed the available flow

$$F_s = \sum_{i\epsilon Is} \sum_{k\epsilon K} F_{si,k} + \sum_{j\epsilon Js} \sum_{k\epsilon K} F_{sj,k} \; ; \qquad \forall\, s\epsilon S \qquad (4)$$

$$F_s\, y_s = \sum_{i\epsilon Is} \sum_{k\epsilon K} F_{si,k}\, y_{s,i} + \sum_{j\epsilon Js} \sum_{k\epsilon K} F_{sj,k}\, y_s \; ; \qquad \forall\, s\epsilon S \qquad (5)$$

The sinks requirements were ensured using total and component balance around sinks k:

$$F_k = \sum_{s\epsilon S} \sum_{i\epsilon Is} F_{,s,i\,,k} + \sum_{s\epsilon S} \sum_{j\epsilon Js} F_{s,j,k}; \qquad \forall\, k\epsilon K \qquad (6)$$

$$F_k\, y_k = \sum_{s\epsilon S} \sum_{i\epsilon Is} F_{,s,i\,,k}\, y_{s,i} + \sum_{s\epsilon S} \sum_{j\epsilon Js} F_{s,j,k}\, y_{s,j}; \qquad \forall\, k\epsilon K \qquad (7)$$

All untreated streams have the same CO_2 concentration as the raw source:

$$y_{s,j} = y_s \qquad \forall\, s\epsilon S \qquad (8)$$

$$y_k = \frac{\sum_{s\in S}\sum_{i\in Is} F,s,i\,,k\; ys,i \; + \sum_{s\in S}\sum_{j\in Js} Fs,j,k\; ys,j}{F_k} \qquad \forall\, k\epsilon S \qquad (9)$$

Any source could be connected to any sink as long as archives the minimum concentration requirement and falls under the capacity limit.

$$F_k \leq G_k^{\max}; \qquad \forall\, k\epsilon K \qquad (10)$$

$$y_k \geq Z_k^{\min} \; ; \qquad \forall\, k\epsilon K \qquad (11)$$

Ideally, all sources can be connected to all sinks. There is no constraint on the number of connections from one source to sink other than the mass balance (max capacity). To avoid unfeasible small connections, and to eliminate fixed costs from the costs correlations, a minimum flow constraint was added using the binary defined below.

$$L_{si,k} X_{si,k} \leq F_{si,k} \leq U_{si,k} X_{si,k} \qquad \forall\, s\epsilon S\ i\epsilon Is\ k\epsilon K \qquad (12)$$

$L_{sj,k}X_{sj,k} \leq F_{sj,k} \leq U_{sj,k}X_{sj,k}$ ∀ sϵS jϵJs kϵK (13)

Where $L_{si,k}$ and $L_{sj,k}$ is the lower flow limit and $U_{si,k}$ and $U_{sj,k}$ is the upper flow limit of treated and untreated sources in pipes respectively. While $X_{s,ik}$ and $X_{sj,k}$ are the binaries associated with flow of the treated and untreated sources respectively. The binaries are designed to have a value (0,1).

To achieve targets set on the city, a constraint was implement on the net capture flow so that

Net capture ≥ Target flow (14)

The target flow is defined by the user whereas the newt capture is defined as follows

Net capture = Total carbon dioxide allocated – Total carbon dioxide emitted (15)

Net capture = $\sum F_k y_{,k} (1-\varepsilon_k - \sum F_{s,i,k} y_{s,i} \varepsilon_T)$ (16)

Non-negativity constraints

$F_{s,i,k} \geq 0$ ∀ sϵS iϵIs kϵK (17)

$F_{s,j,k} \geq 0$ ∀ sϵS jϵJs kϵK (18)

$y_{s,i,k} \geq 0$ ∀ sϵS iϵIs kϵK (19)

$y_{s,j,k} \geq 0$ ∀ sϵS jϵJs kϵK (20)

More specifically the objective function is described by:

Min $TC^{\text{Treated}} + TC^{\text{UnTreated}}$ (21)

Where

$TC^{\text{Treated}} = \sum_{s\epsilon S} \sum_{i\epsilon I} \sum_{k\epsilon K} y_{si} F_{si,k} C^{T}_{si,k} + C^{comp}_{si,k} + C^{Pump}_{si,k} + C^{Pipe}_{si,k} + C^{Pipe}_{s,k} + y_{si} F_{si,k} C^{sink}_{k}$ .....(22)

$$TC^{UnTreated} = \sum_{s\epsilon S} \sum_{j\epsilon J} \sum_{k\epsilon K} C^{comp}{}_{sj,k} + C^{Pump}{}_{sj,k} + C^{Pipe}{}_{sj,k} {}_{s,k} + y_{si} F_{si,k} C^{sink}{}_{k} \quad (23)$$

Where $F_{si,k}\ y_{s,i}$ is the amount of carbon dioxide flow from treated subsource i of source s that is matched to sink k. Likewise, and $F_{sj,k}\ y_{s,j}$ is the amount of carbon dioxide flow of the untreated subsource j of source s that is matched to sink k. $H_{s,k}$ represents the distance between the source and sink processes, $C^{Pipe}{}_{si,k}$ and $C^{Pipe}{}_{sj,k}$ is the pipe cost per unit length of the connection between treated Is and untreated Js source to matched sink (k) respectively based on the calculated diameter and selected standard size accordingly. Since carbon dioxide can be treated and untreated, different type of pipes can be used. $C^{T}{}_{si,k}$ is the cost of treatment applied on the treated source treatment cost as in carbon removal can be used. The untreated source is not associated with a treatment cost, thus in implementation it is considered to be zero. $C^{sink}{}_{k}$ is the costs of the sink processing used, $C^{comp}{}_{si,k}$ and $C^{comp}{}_{sj,k}$ are the total annualized cost of compression to meet sinks requirements. $C^{pump}{}_{si,k}$ and $C^{pump}{}_{sj,k}$ are the total annualized cost of pump to meet sinks requirements. The pumping cost was based on total annual cost obtained through a correlation. All costs were annualized.

The Mixed Integer Non-Linear Program (MINLP) formulation for developed cost-optimal networks have been implemented using “What’s Best 12.0” Lindo Global solver [47] for MS-Excel 2010 via a laptop with Intel Core i7 Duo processor, 8 GB RAM and a 32-bit operating System. The implementation contained 169 constrains, 48 binaries and 50 continuous variables.

7. CASE STUDY AND IMPLEMENTATION

The purpose of this section is to apply the approach to an illustrative case study. Open literature data were used to construct an industrial city to demonstrate the applicability of the proposed method. Using carbon dioxide emission data and choosing several representative carbon producing sources; the following set up was investigated.

7.1 Step 1: Data acquisition for industrial zone

The city is assumed to comprise of several plants, with a cumulative carbon footprint of 20,464 t/d. This corresponds to around 7.5 million tons carbon dioxide per year. Table 2 below shows the production capacity data for each of the plants that have been considered in this study. Five plants in total were chosen to represent a small city consisting of ammonia, iron and steel, refinery, power plant, and fuel additive productions. All plants involve a diverse mix of carbon dioxide emitting sources. Their corresponding total carbon dioxide emissions are all listed and detailed in Appendix A.

Table 2: Industrial City Plants and Produced Carbon Dioxide

Plant	Capacity	Produced CO_2, MTPD
Fertilizer Complex	1497 MTPD Ammonia	
	1601 MTPD Urea*	3315
Iron and Steel Production	2 MTPD	5012
Power Plant	1 GW	9385
Refinery	66,400 bpd of Crude Oil	2184
Fuel Additive Plant	2500 MTPD Methanol*	
	1850 MTPD MTBE	568
Total Emission		20,464 MTPD

*Urea and methanol are considered sink processes

Theoretically, all of the CO_2 emitting processes within the plants listed above can be included in the analysis. However, this case study considers four main CO_2 emitting sources, representing about 70% of the industrial zone were chosen for carbon integration. CO_2 emissions that were considered in this study were assumed to come from fertilizer production, iron and steel, an oil refinery and natural gas fired power plant. A representative stationary source from each of the sources that have been chosen ensures varied compositions and flow rates within the city while representing the majority of its emissions. Source data collected, provided in Table 3 below, were assumed to be at a temperature of 298K and atmospheric pressure of 101 kPa. It should be noted that for simplicity reasons, multiple CO_2 emitting sources within the same plant were lumped into a single stream of the same composition. Appendix A details the assumptions and references used in identifying compositions, flows and emission points.

Table 3: Carbon Dioxide Chosen Sources

Source	wt% dry	mol% dry	vol% dry	ρ dry, kg/m^3)	Estimated C_{si}^T (USD/t CO_2)	CO_2, MTPD
Ammonia-CO_2 amine unit	100%	100%	93%	1.85	0	977
Steel-Iron	44%	20%	21%	0.97	29	3451
Power-gas turbine	7%	3%	3%	0.74	43.15	9385
Refinery-boiler	27%	11%	11%	0.81	34.8	1092

7.2 Step 2: Identification of carbon sinks

In addition to isolating all carbon dioxide sources utilized in this study, several processes were also identified as major CO_2 sinks within the city. The fertilizer complex includes a urea producing facility that primarily consumes most of the ammonia and carbon dioxide produced within the complex. However, excess ammonia that does not react due to carbon dioxide limitations and is sold as by-product. This creates a sink for excess carbon dioxide within the city, which can react with excess ammonia to produce more urea. Carbon dioxide can also be consumed as a carbon building block replacement to conventional fossil-driven processes. For instance methanol production, which is one of the city's plants, operates using the conventional syngas technology. Natural gas driven syngas can be replaced by a mix of carbon dioxide and hydrogen, both can be available within the city as by-products. This process of methanol production has been demonstrated on pilot scale, using renewable energy sources (Olah et al, 2008[48]). Hence, a methanol production facility was assumed to be present as a potential carbon dioxide sink.

Other potential sinks include both biological and geological sequestration. Biological sequestration, such as carbon dioxide use in algae production and horticulture greenhouses, can also be applied to generate revenue and fix CO_2. Moreover, storage of carbon dioxide in saline formations can also be considered as a possible sink to mitigate city produced emissions. In addition, since Qatar's oil reserves are gradually depleting, the application of enhanced oil recovery, also a potential carbon dioxide sink option, can be beneficial. The capacity of the sinks, purity requirements, and costs are all

summarized in Table 4. Appendix B further givens the references and assumptions used to obtain sink specifications.

Table 4: Sinks Considered in Carbon Integration

Sinks	CO_2 Composition. (wt%)	Flow CO_2, MTPD	P, kPa	$C^{sink}_{k,}$ USD/ton CO_2	ε k
Algae within City	6	283	101	0	0.42
Greenhouses out of City	94	1030	101	-5	0.5
Methanol Solar	99.9	1710	8080	-21	0.098
Urea	99.9	1126	14140	-15	0.29
CO_2-EOR	94	8317	15198	-30	0
Saline Storage	94	8317	15198	8.6	0

7.3 Step 3: Treatment, transmission cost data

This step explains the costs associated with the system. In the case study, a total of 8760 operating hours per year were assumed for all plants, and the entire city. The cost of electricity was at 0.02USD/kWh, while all capital costs were annualized using a capital recovery factor (CRF). The CRF has been calculated using the following equation:

$$(CRF) = \frac{i(1+i)^n}{(1+i)^n-1} \quad (24)$$

The interest rate, i, was assumed to be 8% while the n was taken as 10, thus the CRF was estimated to be 0.15.

Regarding treatment, there exist several technologies that can concentrate and remove carbon dioxide from a given stream. These technologies are often called ‘carbon capture’. In this work, such technologies are referred to as ‘removal technologies’ to be

distinguished from the concept of capturing CO_2 in utilization sinks. Carbon dioxide removal technologies can be one of the following: oxy combustion, pre-combustion removal and post combustion removal (Wong and Bioletti, 2002[49]). Post-combustion removal was assumed to be the only treatment technology applied for all sources with the city studied. More specifically, chemical adsorption using amine solvents was applied to its relative easy application and level of maturity. Moreover, amine solvents have a great ability to capture low concentrations of carbon dioxide (Chapel et al, 1999[50]). Therefore, since the aim of this work was to ultimately reduce the carbon dioxide footprint as much as possible, chemical adsorption using amine solvents was assumed for treatment. Due to the large flows of CO_2, each treatment unit was assumed to be associated with one source. The cost parameters were estimated based on available literature, excluding the cost of compression and delivery, specific to each source. The treatment cost of each source has been included in Table 3. This is due to the different nature of each source that would require its own costing method. The appendix shows the calculation method that has been adopted, for all sources.

The next step in establishing a carbon integration connection is to account for the cost of compression. Compression is needed in both overcoming pressure drop in pipelines and to meet sinks pressure requirements. The cost of compression is usually dependent on power requirement if a compressor is used or on volumetric flowrate if a blower is applied. Since pure CO_2 is produced in treated sources, the treated sources were compressed isothermally using a centrifugal compressor based on its power requirement. Based on the assumption of known compositions, total flow, pressure drop

in pipes and final pressure, the specific power for each connection was calculated in advanced using Aspen Plus simulation. Appendix shows the specific power for each connection. The capital cost of compression is calculated using the correlation below

$$CC^{capital}{}_{treated} = 158{,}902\ (\frac{\text{capacity of current comp.}}{224\ \text{kW}})^{0.84} CRF \quad (25)$$

The cost of compression was different for the untreated sources. This is mainly due to the large volumetric flow rate of flue gas and its composition that might affect the equipment performance. Thus, a blower was used to calculate the compression based on the volumetric flow. The compressor capital cost is estimated as the above treated correlation, while the blower was based on a known cost with a known capacity raised to a specific exponent. This value was brought to current cost using index values and was adopted based on Durão [51] However, since the maximum pressure that can be reached using a blower is far less than that of an industrial compressor, it is assumed that a compressor is used to reach the final sink pressure similar to the treated source. The switch between compression devices is assumed to occur at 5000 kPa (Towler and Sinnott, 2007[52]).

$$CC^{capital}{}_{untreated} = 79{,}998\ (\frac{\text{capacity of current comp.}}{473\frac{\text{m3}}{\text{s}}})^{0.6} \quad (26)$$

The operating cost for both sources is calculated by

$$CC^{operating} = power\ *\ Elec.(\frac{USD}{kWh}) * \left(\frac{days}{year}\right) * 24\ h \quad (27)$$

The overall cost is the summation of the capital and operating cost

$$C_{comp} = CC^{capital} + CC^{operating} \quad (28)$$

To avoid two-phase flow within a pipe that could damage equipment and result in inefficient delivery, a pump is installed to move the fluid once it passes the critical point. The maximum pressure to sustain the gaseous state was 7.38 MPa after which the CO_2 stream would be at critical condition based on McCollum and Ogden [53] assumptions. The cost of the pump was calculated based on the flow and power requirement for both treated and untreated sources using the same correlation

For treated sources:

$$\Delta P_{si,k} + \Delta P_{si,k}{}^{pipe} > 7.38 \text{ MPa} \tag{29}$$

$$W_{si}{}^{p} = \frac{1000*10}{24*36}\left[\frac{F_{si}(\Delta P_{si,k}+\Delta P_{si,k}{}^{pipe}-7.38*10\text{^}3Pa)}{\rho\eta}\right] \tag{30}$$

$$C^{pump}{}_{si,k} = [(\,1.11 * 10^{6}\frac{W_{si}{}^{p}}{1000} + 0.07 * 10^{6}) * CRF\,)\,] \tag{31}$$

For untreated sources:

$$\Delta P_{sj,k} + \Delta P_{sj,k}{}^{pipe} > 7.38 \text{ MPa} \tag{32}$$

$$W_{sj}{}^{p} = \frac{1000*10}{24*36}\left[\frac{F_{sj}(\Delta P_{sj,k}+\Delta P_{j,k}{}^{pipe}-7.38*10\text{^}3Pa)}{\rho\eta}\right] \tag{33}$$

$$C^{pump}{}_{sj,k} = [(\,1.11 * 10^{6}\frac{W_{sj}{}^{p}}{1000} + 0.07 * 10^{6}) * CRF\,)\,] \tag{34}$$

Carbon dioxide can be transported in liquid form in tanks via trucks, larger quantities by ships and pipelines. Since the problem is based on an industrial city, onshore pipelines were the only mode of transportation considered. The pipeline cost was calculated based on a linearized pipeline data of hydrogen transportation presented by Parker [54]. A linear correlation was produced using MS Excel regression in terms of

USD/mi in 2000, which converted to 2009. The same correlation was applied to both treated and untreated sources.

$$C_{si,k}^{\text{Pipe}}\left(\frac{\text{USD\$}}{\text{mi}}\right) = [50193\ (D^{c}{}_{si,k}) + 51075] * (748.1/394.3) * \text{CRF} \quad (35)$$

$$C_{sj,k}^{\text{Pipe}}\left(\frac{\text{USD\$}}{\text{mi}}\right) = [50193\ (D^{c}{}_{sj,k}) + 51075] * (748.1/394.3) * \text{CRF} \quad (36)$$

The diameter of the pipe was calculated based on based on given mass flow rate, known temperatures, streams molecular weight, pipe pressure drop and assumed outlet velocity from the sources. The velocity throughout was assumed to be 20 m/s, while the molecular weight was estimated based on the different streams compositions.

$$D_{si,k} = \sqrt{\frac{8.314 TsiF_{si,k}}{v_{sik} M_{si}[\Delta P_{s,k} + \Delta P_{s,k}{}^{pipe}]} * \frac{4}{\pi}} \quad (37)$$

$$D_{sj,k} = \sqrt{\frac{8.314 T_{sj} F_{s,k}}{v_{sjk} M_{sj}[\Delta P_{s,k} + \Delta P_{s,k}{}^{pipe}]} * \frac{4}{\pi}} \quad (38)$$

Pipelines are often available in standard sizing; therefore, the standard size was obtained by rounding up the calculated diameter

$$D^{c}{}_{si,k} = Round\_Up(D_{s,ik}) \quad (39)$$

$$D^{c}{}_{sj,k} = Round\_Up(D_{s,jk}) \quad (40)$$

Standard commercial sizes were used up to 42-inch diameter pipe. Larger flows are assumed to require the installation of multiple pipes in parallel. The distances between all sources and sinks were obtained assuming the shortest distance on a given layout shown in Table 5. The pressure drop in pipelines is directly used in the diameter calculation and is often obtained via a trial and error method involving the calculation of

the friction factor. Here however, a heuristic was assumed to obtain the approximate pressure drop based on Coker [55], the pressure drop was calculated as follows

$$\Delta P^{pipe} = \frac{0.1\ bar}{100\ m} * \left(\frac{\mathrm{v}\left(\frac{\mathrm{m}}{\mathrm{s}}\right)}{61\left(\frac{m}{s}\right)} \right) * \frac{100\ kPa}{1\ bar} * \mathrm{Hs,k} \quad (41)$$

Table 5: Distances Between Sources and Sinks (km)

Source/sink	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia	1.72	25.38	1.51	1.51	1.55	1.56
Steel	2.07	25.73	1.86	1.86	1.9	1.91
Power plant	2.77	27.33	2.95	2.95	0.91	0.51
Refinery	2.53	27.09	2.71	2.71	0.66	0.82

7.4 Step 4: Identification and design of cost effective integration networks

The model was run with the objective of achieving the lowest total cost network. The total cost for each target was calculated, which includes the cost of carbon removal, the cost of compression, and the cost of transportation i.e. piping and delivery. The optimization was solved for a set of targets, 10%, 30%, 40%, 50%, 50% being the maximum target the optimization can reach based on inefficiencies and carbon losses from sinks and treatment. A number of intermediate points were chosen to understand the flow and type of connections. While the carbon removal, compression and transportation added to the total cost it can be observed that the network is able to obtain positive revenue especially in the case of smaller flows.

For a target of 3.0% from the overall city, the main connection obtained was from the purest treated source i.e. ammonia carbon removal process to the highest

revenue sink, EOR. The total cost of the connection was -588,000 USD per year. The compression cost being the highest. The total cost breakdowns as well as the connections are reported in Figure 6.

Enhanced Oil Recovery continued to be the selected when the capture target was increased 10% and gave a total cost of 1.17 million USD per year as shown in Figure 7. The number of connections increased to include two sources, ammonia amine removal unit as previously observed and steel iron mill source as the second connection. The second source used as was the next purest however, no profit was gained that was attributed to the increased flow rate.

When the capture target was increased to 30%, the power plant source was used as the third connection established to supply the remaining carbon dioxide demand to the enhanced oil recovery sink instead of using the refinery's more concentrated and less expensive supply, which was connected to solar methanol production instead. Shown on Figure 8, the selection of sources is explained due to the compression and delivery cost in addition to the flow rate demand of EOR. Based on revenue, it would have been expected that the urea would the second selected source after EOR. However, the optimization preferred solar methanol sink despite the sinks having the same concentration requirement, solar methanol was chosen. This is due to pressure and sinks efficiency. Lower pressure requirement needed at the methanol sink than at urea, thus less compression cost. Moreover, solar methanol sink has higher sink efficiency, releasing less carbon dioxide than urea thus storing CO_2 longer and costing less overall to capture a CO_2.

Interesting observation was noticed at 36% capture, Figure 9, corresponding to 7,500 t/d net carbon dioxide. The number of connections increased to 6 connections. Ammonia carbon dioxide removal unit was connected to EOR along with treated steel and treated power plant flows making the concentration at sink 100 wt%. The remainder of the steel plant flow as untreated was directed to the algae sink. While part of the power plant flow was split between storage and solar ammonia. Storage is the highest sink, yet it was chosen over other revenue generating sinks such as Urea and Greenhouse. The connection to storage would require the supplier to pay 8.6 USD per ton of carbon dioxide. However, due to the high sink efficiency, storage is far less likely to leak carbon dioxide than the other sinks with the exception of enhanced oil recovery. Thus, the optimal solution prefers longer sequestration than short-term capture that initially seems less expensive, but is more costly on the long run.

The connection of untreated steel at 44 wt% carbon dioxide to algae, which required a minimum of 6% rather than connecting the less valuable power plant at 7% to algae, is attributed to the large total volume of carbon dioxide that would be transferred from power to algae. Transferring untreated low concentration flow from the power plant to the algae sink would have to overcome pressure drop through compression, to compress and deliver steel at a higher concentration seems to be cheaper as determined by the optimization. This agrees the general pattern with small flow compression cost, as smaller flows cost more to compress than larger flows as reported by Hasan et al [41] noticed in their network of carbon-sink matching. The refinery source was not chosen at all in this allocation.

The slight increase of 4% more capture to make an overall 40% capture of city, keeps the same number of connections with switching some of the previously made decisions as shown in Figure 10. Interestingly, the carbon capture network abandons the algae sink as the target is set. At 40% city capture, corresponding to 8185 t/dCO_2, ammonia removal was connected to produce solar methanol. The steel carbon dioxide stream was fully connected to EOR along with the power plant as treated sources. While some flow from power plant was connected to storage. The overall network at this target cost was 24 USD million per year. The optimization continued to prefer the longer storage i.e. the more carbon capture efficient sinks as the lowest cost solution rather than rely on the profit solely. The choice of sources switches from the last target where ammonia removal is chosen to supply the solar methanol along with treated refinery and steel. While, steel and power both treated are chosen to fill the EOR sink.

Solving the model for max carbon dioxide use, it was calculated that 10,291 t/d CO_2 corresponding to round 50% of the total city emissions was the limit to attain a feasible solution. This is due to mass limitation, as all of the carbon sources were exhausted in the carbon integration. Losses of carbon were large between sink inefficiency and treatment. At 50% capture of the city emissions, shown in Figure 11, the connection between ammonia carbon dioxide removal and solar methanol switched from last target from solar methanol production as a sink to EOR. The refinery source was connected fully as treated flow to EOR while, the steel source was split between storage and EOR. Untreated steel was mixed with treated flow from power plant at the storage sink making the composition at storage 94 wt%.

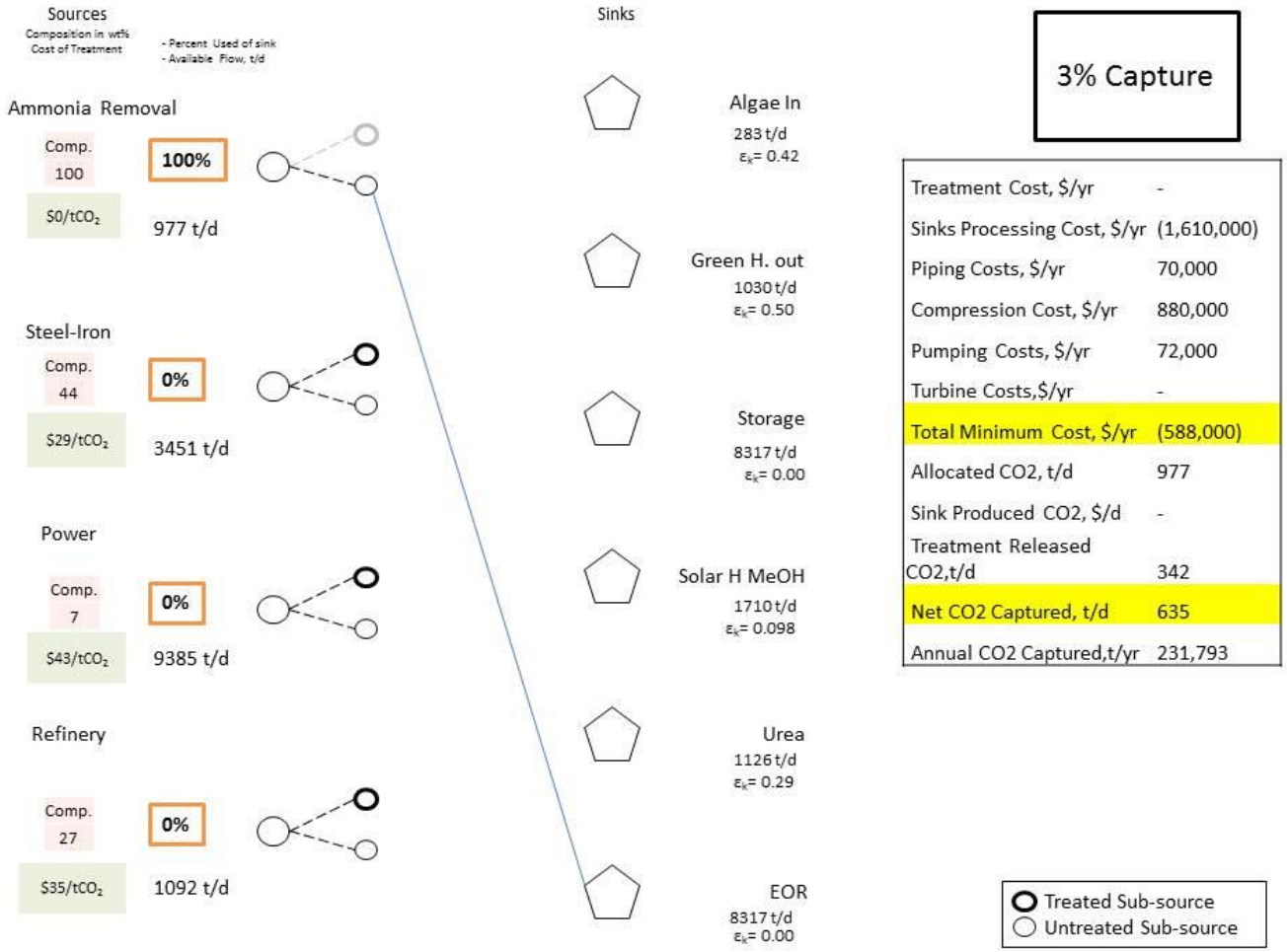

Figure 6: 3% Capture

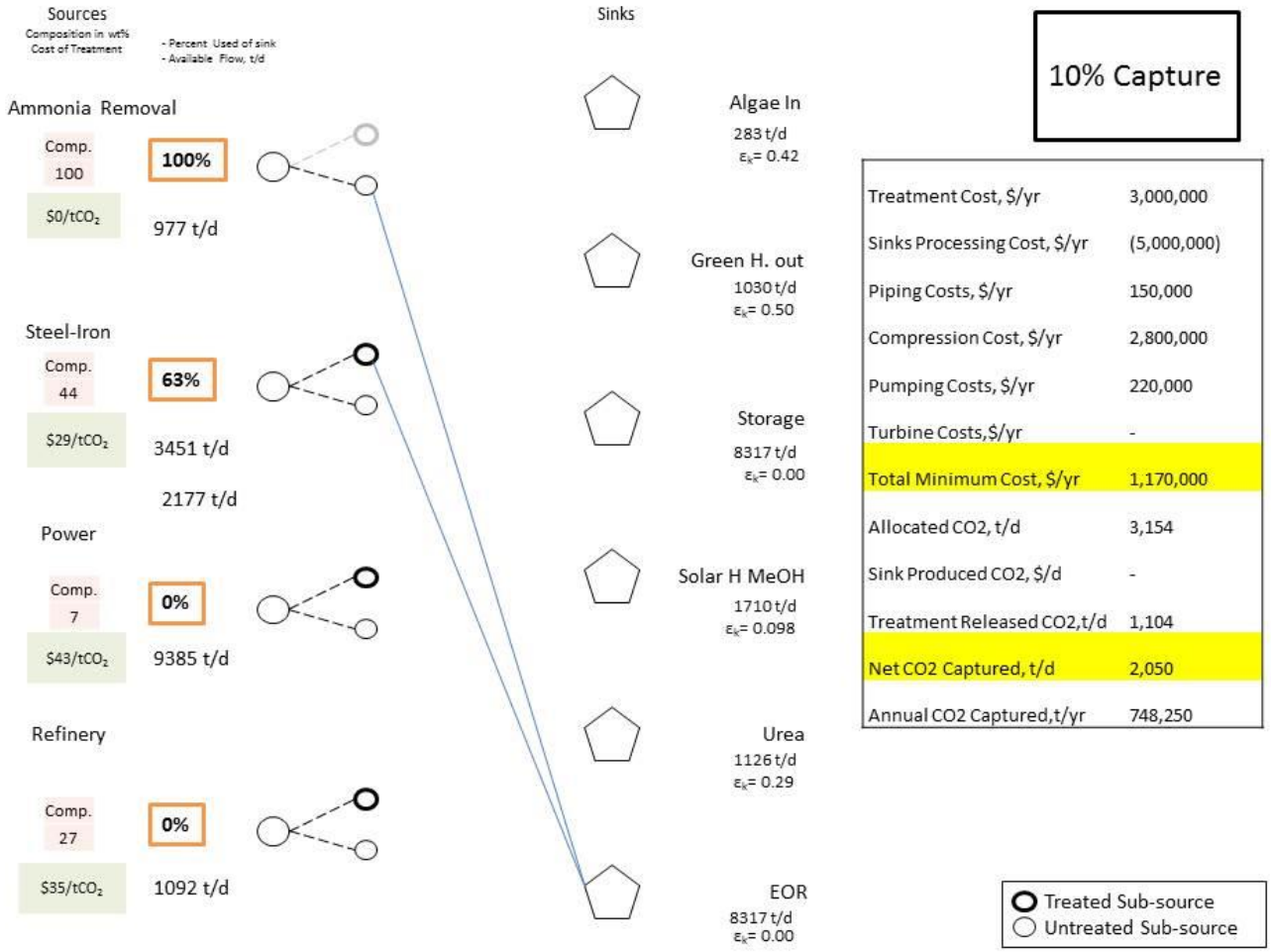

Figure 7: 10% Capture Target

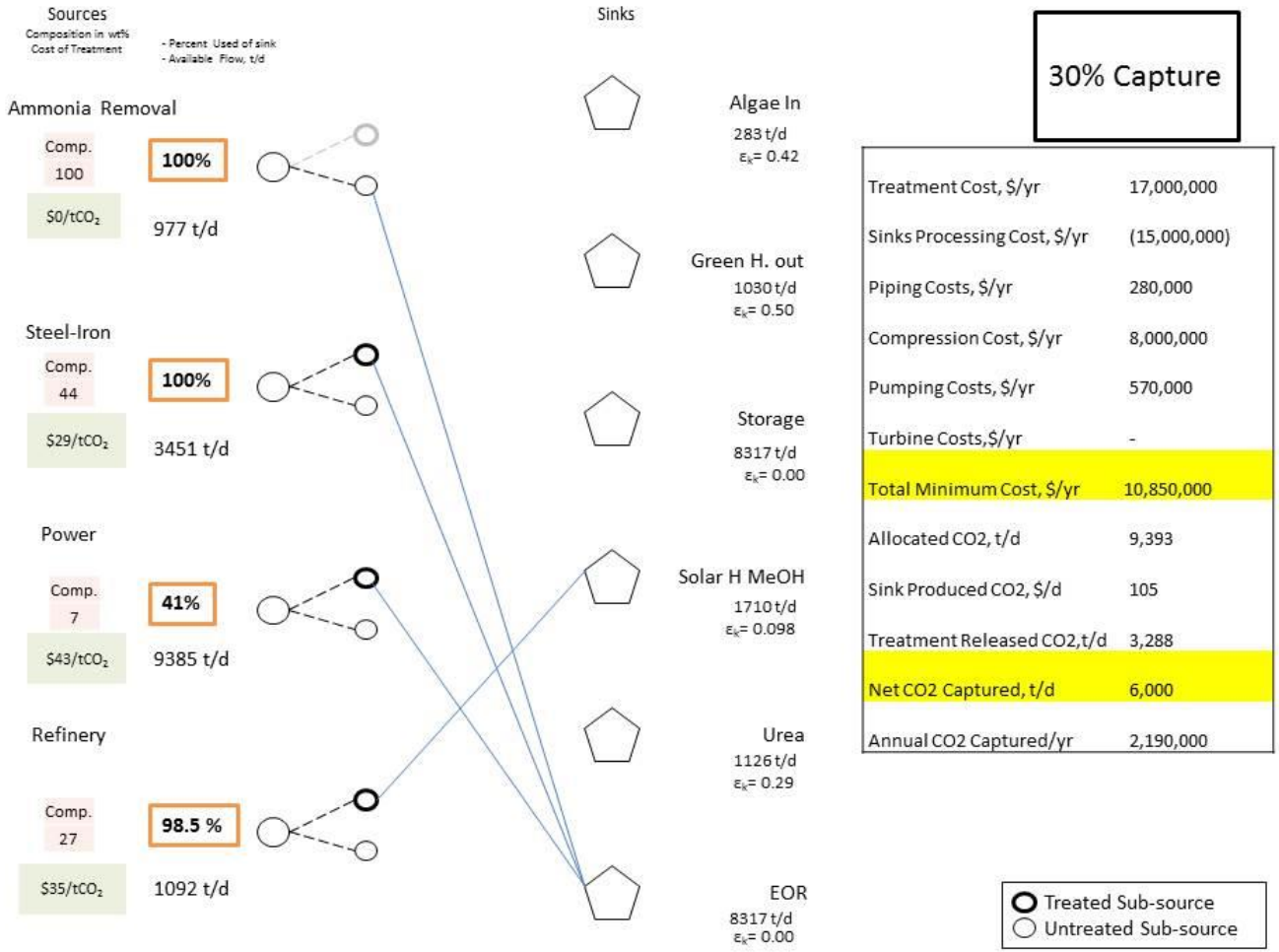

Figure 8: 30% Capture Target

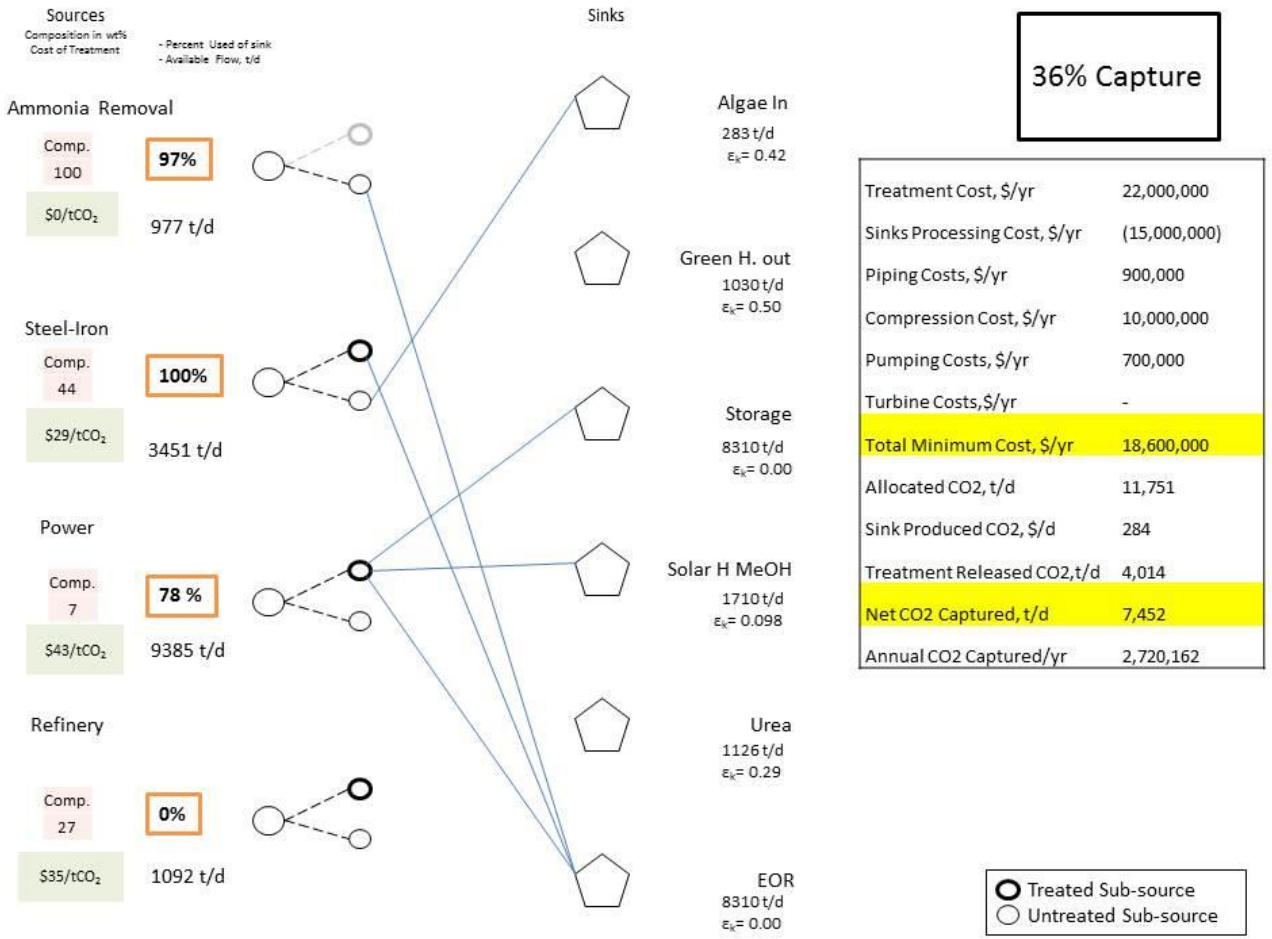

Figure 9: 36% Captured

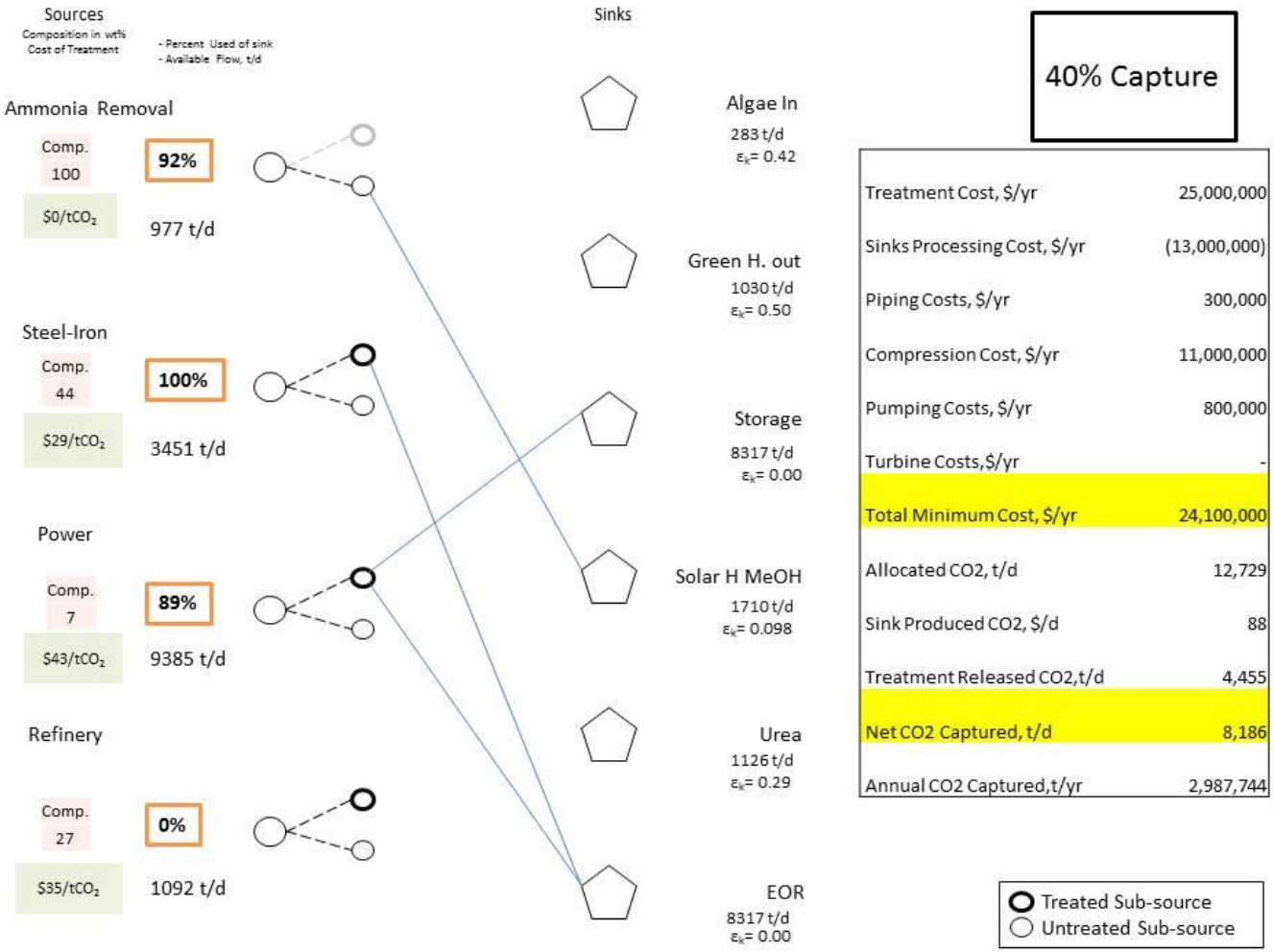

Figure 10: 40% Capture Target

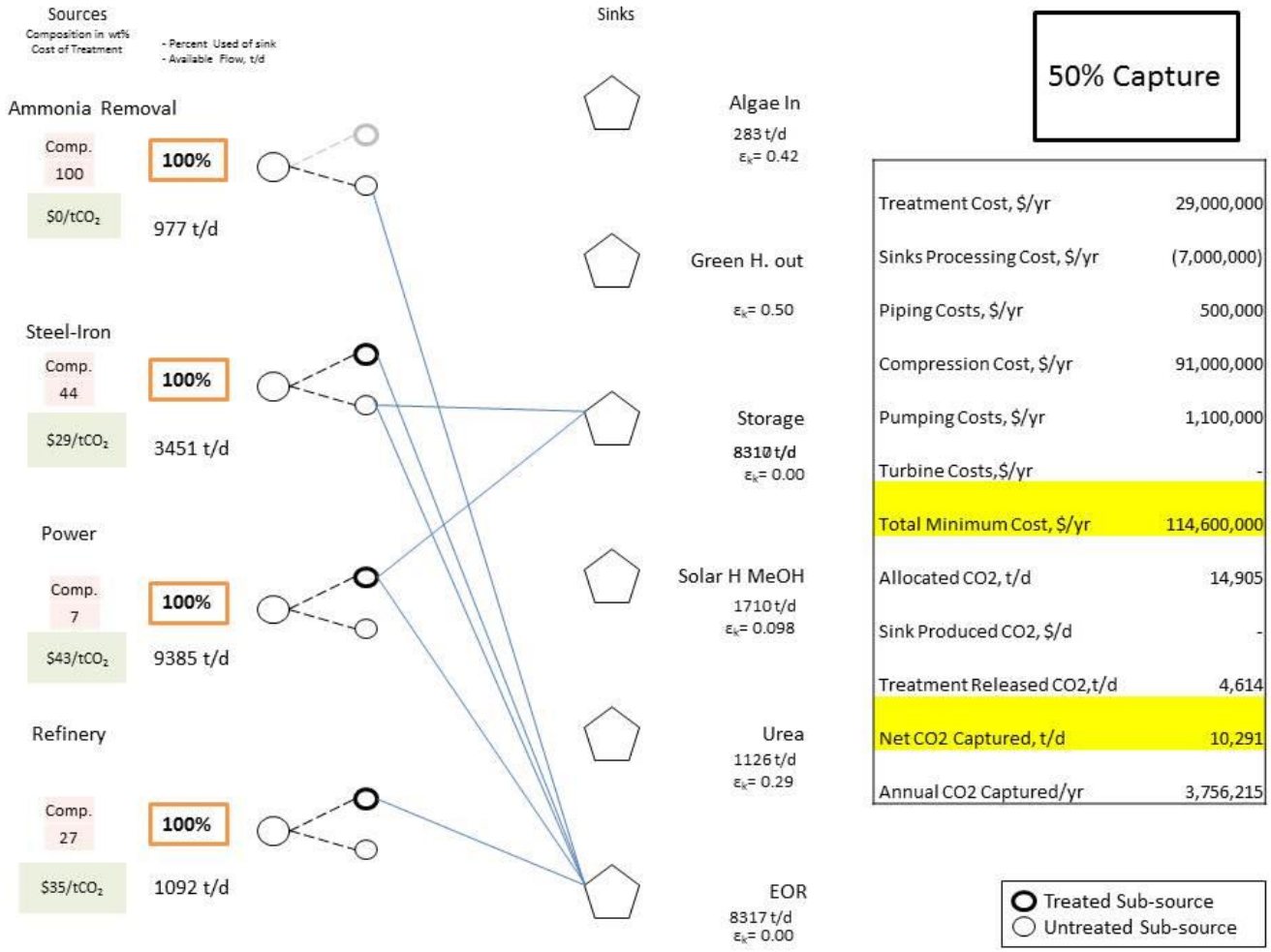

Figure 11: 50% Capture Target

The overall network at this target cost 114 USD million per year, extremely higher that the capture trend as shown in Figure 12. The major cost element contributing to the increase comes through compression cost. This significant increase happened at before the 50% capture point. Looking at the 49% capture explained the connections logic. At 49% capture, shown in Figure 13, all the sources flows are exhausted completely used at 100% each. The difference in the cost comes from the connections at sinks. Ammonia removal source was connected to the solar methanol sink and to storage. Due to sink efficiency, more carbon dioxide is captured at storage and EOR. This explains the shift of the ammonia removal connection from solar methanol at 49% capture to EOR at 50% capture. As more cuts of carbon dioxide were required by the set target, steel was transported as a dilute flow at 50% capture, when at 49% was transported as a concentrated flow. This switch saves the amount of carbon dioxide that would have been released in through the treatment unit as explained by the carbon removal efficiency factor in Appendix A. Therefore, transporting untreated dilute carbon dioxide increased the compression cost significantly and achieved minor carbon savings.

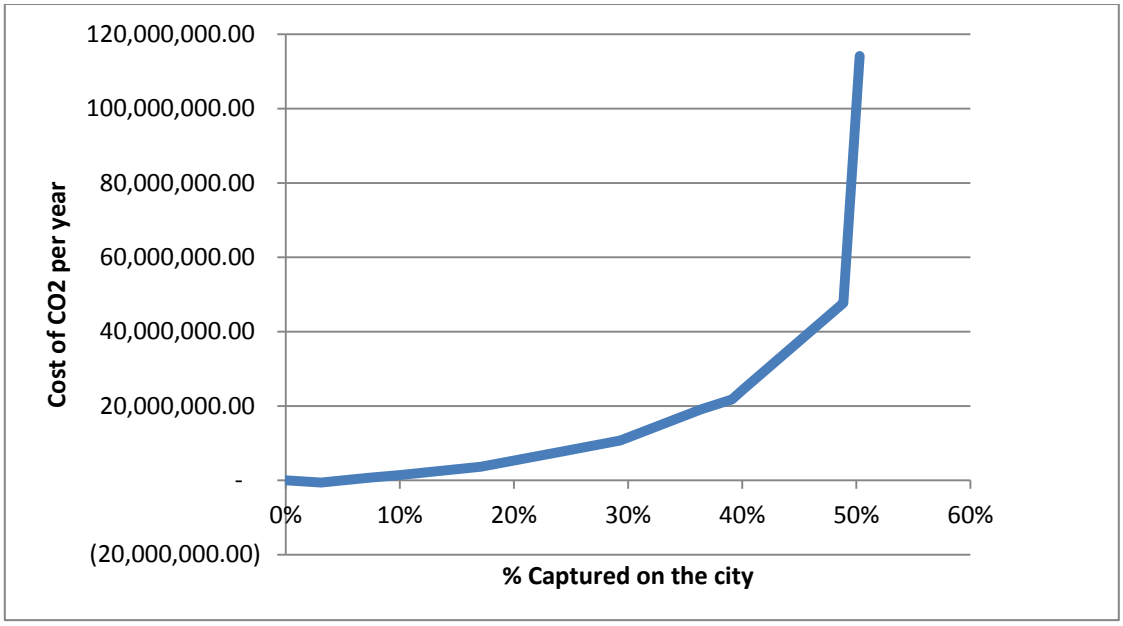

Figure 12: Minimal cost carbon reduction network

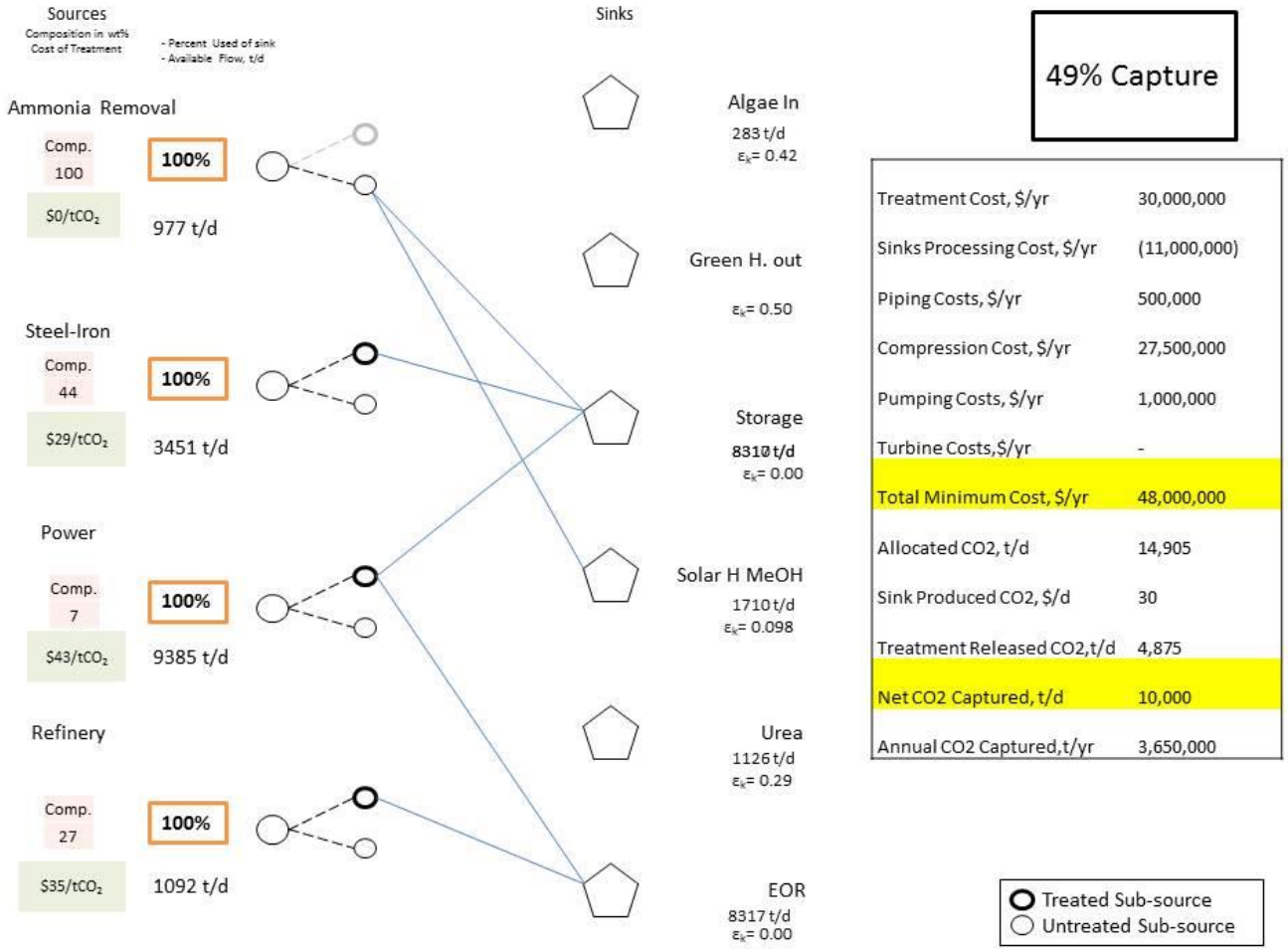

Figure 13: 49% Captured

The overall optimal carbon reduction network that designed to manage 70% of the total stationary emissions from the industrial city results is shown in Figure 14. The general trend followed was that the purest sources, having less treatment cost were chosen first and allocated to the highest revenue sink. The results showed an exponential relationship between total cost and capture rate as we moved closer to the maximum target Figure 12. As higher reduction targets are set, the network moves towards carbon savings at sinks and at more extreme cuts through treatment options. This indicates that in order to cut more carbon dioxide in a cost effective manner, larger flows should be used by the addition of more sources.

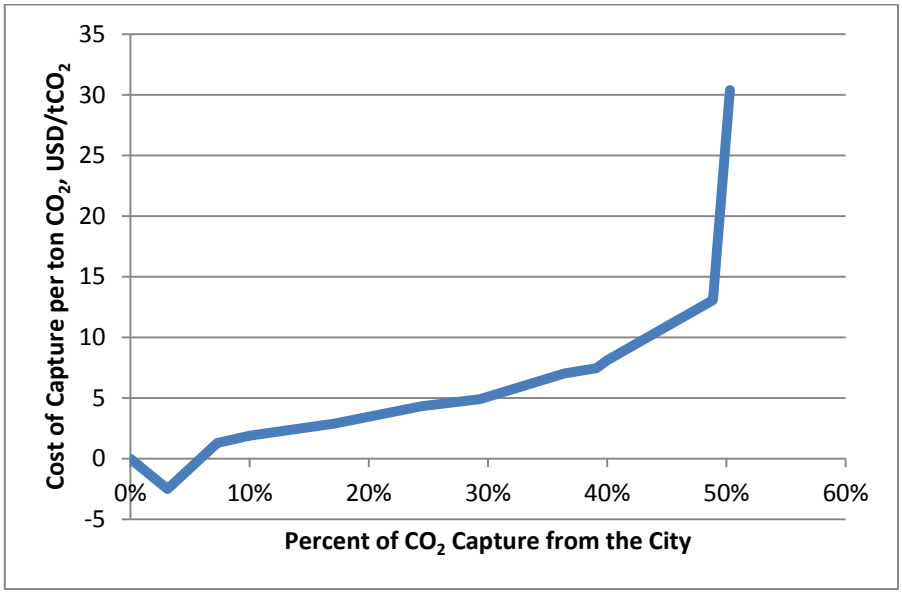

Figure 14: Carbon Integration Cost per ton vs Capture

The cost of carbon integration network is considered relatively lower than expected. Several factors contribute to obtain this cost-effective network. At smaller flows, it mainly has to do with the price of carbon dioxide in today's economy. Enhanced oil recovery was the main chosen sink filled with the relatively pure sources, the cost of treatment and delivery was overcome by revenue as high as 30 USD per ton CO_2. The short distances in an industrial city as well as the diverse sources helped reduce some of the preparation costs. Thus, the method can be effective if a profitable sink was present in an industrial city.

For large emission cuts the cost optimal sinks are the ones with large storage and longer storage time. This means that CCS would continue to be the preferred sink for larger reduction target. Compared to other sinks, storage and enhanced oil recovery can process more carbon dioxide, while carbon release is minimal. Thus, inefficiencies of sinks and treatment play a deciding factor for the sinks selection. Improving carbon processing technologies efficiencies and capacities would be interesting to test whether or not it can compete with geological storage.

8. CONCLUSION AND FUTURE WORKS RECOMMENDATIONS

This work has presented an overall methodology for carbon integration for footprint reduction in industrial zones through utilization; a representation of the carbon integration problem and an optimization approach to explore the representation in order to identify low cost options for given carbon reduction goals. It is a systematic and generally applicable tool for the optimal design of the problem carbon footprint reduction across multiple processes scale and a step towards mathematically modeling possible synergies of eco-industrial parks. The work as a reduction method on an industrial city scale has given the following advantages:

- Develop a framework that considers different components needed to create a network.
- Applicable as a tool that can help engineers and industrial designers in generating and screening processes within a city.
- Applies an easy to use step-wise method that excludes non-attractive potential sinks and candidates based on a set of criteria prior to any calculations.
- Accounted for transportation, delivery and preparation mechanisms and costs.
- Incorporated market demand and available resources in the selection step.
- Can assist policy makers in gaining valuable insights on the achievable target limits, with the aim of developing national emission cuts using various existing or potential technologies.

- Optimization of the system using a non-linear program that can be adopted onto various scenarios.
- Incorporate off-setting carbon removal (carbon capture) cost by generation of value added product from waste within an industrial city
- Applied a fundamental concept of IE by creating synergies by use of a waste product as a raw input into another process

An optimization-based approach has been introduced for the systematic design and integration of carbon dioxide in an industrial city. A modified source-sink structural representation has been developed to include various configurations including materials exchange, stream splitting or treatment in preparation step, and allocation to sinks. An optimization formulation based on a mixed-integer-non-linear program has been developed to represent the concept. A case study has been generated and solved to show the effectiveness of the developed approach and the optimization model formulated, as well as to illustrate the method applicability on real systems.

Recommendations for future work include

1. Re-visiting the formulation to allow mixing at source in addition to mixing at sinks to investigate effect on network and compression.
2. Temporal aspects of the problem would be vital in order to develop integration over time.
3. The use of different carbon removal (i.e. capture) processes based on compositions and advanced technology.

4. The use of renewables to replace some of the carbon dioxide produced as opposed to allocating CO_2 in sinks and expanding the method to consider renewable energy options and fuel switching.
5. Incorporation of heat and water as both are contributing factors.
6. The variation of pressures required as part of approach and exploring power recovery and consumption.
7. The introduction of new components into the allocation approach (e.g. hydrogen) and exploring their interactions.
8. The exploration of economic studies based on multi-share companies and tracing costs.

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APPENDIX A

Appendix A. Source Data Gathering and industrial Streams

A1. Stream Data

Table 6: Stream Data of Industrial City

Plant	Source	Flow	% of City
Refinery	Boiler	1092	5%
Refinery	Utilities	437	2%
Refinery	Fluid Catalytic Cracker (FCC)	437	2%
Refinery	Hydrogen production	218	1%
Steel	Iron production	3451	17%
Steel	EAF off-gas	1154	6%
Steel	Mini mill Electric Arc Furnace (EAF)	407	2%
Ammonia	Excess CO_2 out of amine unit	977	5%
Ammonia	Primary reformer	2338	11%
Methanol	Emission	568	3%
Power plant	Gas turbine	9385	46%
	Total Emissions	20,464	

A2. Steel Plant

Steel plant emission points and carbon dioxide flow was estimated based on a values obtained from Ho et al [56].

Table 7: Steel Production Data

Description	Iron production (1.5 Mt/yr) Source 1: Midrex stack gas	Steel production (1.5 Mt/yr) Source 2: Midrex EAF off gas	Steel production (0.5 Mt/yr) Source 3: Mini mill EAF off gas
Pressure (kPa)	101.3	101.3	101.3
Temperature (°K)	706	573	573
Total mass flow rate (kg/s)	109.08	25.8	9.1
Total mass flow rate, MTPD	9424.9	2228.84	786.65
Velocity (m/s)	20	20	20
Flue gas composition (wt%)			
N_2	44.66	46.12	46.12
H_2O	16.85	0.53	0.53
CO_2	36.62	51.76	51.76
O_2	1.87	1.59	1.59
CO_2 flow, MTPD	3451.256294	1153.792512	406.957824

A3. Fertilizer Plant Complex

The main emission points from the fertilizer plant are assumed to be primary reformer and amine removal unit. Ammonia is produced using natural gas syngas via air blown primary and secondary reformers as shown in the figure below. The process parameters are estimated based on European Commission Best Available Technology [58].

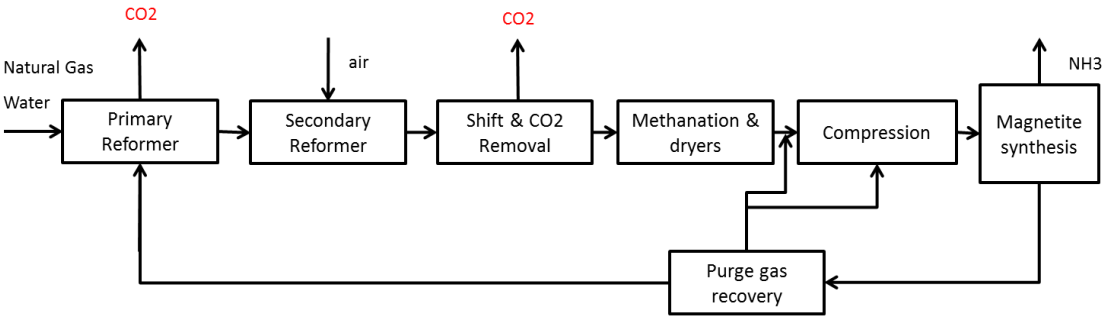

Figure 15: Ammonia Production (Strait, 2010[57])

In the primary reformer, conventional catalytic steam reforming of natural gas takes place according to the following reaction:

$CH_4+H_2O \rightarrow CO+3H_2$

Steam is fed at molar ratio of 3 steam to carbon and part of the inlet natural gas is burned to supply heat for the endothermic reaction to take place. The resulting flue gas contains 8% vol carbon dioxide. At a ratio of 500 kg CO_2 produced for GJ natural gas, and assuming the process consumes 31 GJ per tons ammonia, 2338 tons per day CO_2 is produced as flue gas [58].

It was assumed that carbon dioxide produce from the steam/air reforming step was at 1.18 kg/kg NH3. Assuming that the produced amount of carbon dioxide is completely recovered from the amine unit since the treated gas is left with less than 50 ppm and since the complex would consume 1.29 tCO2/tNH3 to produce urea [58] this leaves an excess of 977 MTPD of pure CO_2 emitted. The following table summarizes the balance involved.

Table 8: Fertilizer Complex Data

Calculation	Amount	Unit	References/Notes
Ammonia production	1497	MTPD	(Daghash and Benyahia,2014 [59]) (80% of ammonia to convert to urea)
Carbon dioxide produced	1766.46	MTPD	EU Best Available Technology [60]
Amount of NH_3 used in urea	611.8239	MTPD	[59]
Amount of CO_2 used in urea	789.252831	MTPD	[58]
Urea produced	1061.4	MTPD	[58]
Excess ammonia	885.1761	MTPD	Balance
Capacity urea	1536.8	MTPD	Calculated based on [58] ratio
Capacity CO_2	1126.45	MTPD	
Electricity	32425.72	kWh	
Energy for electricity (natural gas)	3664.11	m3/day	Gas turbine exhaust (urea utilities)
Actual energy	11103.35	m3/day	33% efficient
Excess carbon dioxide from amine	977	MTPD	Balance

A4. Refinery

Refinery produced carbon dioxide account for around 4% of the global carbon dioxide emissions; this means that 1 billion tons of carbon dioxide per year are released to the atmosphere. The refining sector ranks third after power generation in terms of carbon dioxide emission. Most of the carbon dioxide generated comes of the use of 1.5 to 8% of the refinery raw material as fuel in order to supply the required heat. Carbon dioxide from a refinery complex can be emitted from the following processes as described by Van Straelen et al, [60] .Based on [60] emissions from a refinery are as mentioned in Table 9.

Table 9: Refinery Emission Breakdown

Source	Emissions
Furnace and boilers	30-60%
Utilities	20-50%
FCC	20-35%
Hydrogen manufacturing	5-20%

Assuming a plant capacity of 66,400 bl/d of crude oil and making an approximation based on the emissions reported above, the following emissions are obtained.

Table 10: Refinery Emission Points

Source	Percentage of emissions %	Flow, t/d CO_2
Boiler	50	1092
Utilities	20	436.8
FCC	20	436.8
Hydrogen	10	218.4
Total	100%	2184

A5. Power Plant

Power is supplied to the city using a 1 GW natural gas combined cycle (NGCC) power plant. The net efficiency is assumed to be 49.62% and with a carbon footprint of 0.366 kg CO_2/kWh (Karimi et al, 2012, [61]). Main emissions come from the gas turbine exhaust, which is used to generate steam through Heat Recovery Steam Generator (HRSG). The flue gas was estimated to be 9385 tCO_2/d. Flue gas composition was given in volumetric percent by Jordal, K.[62], et al, as shown below and converted to mass fractions.

Table 11: Flue Gas Compositions of Power Plant

Component	Composition vol%	Composition wt%
N_2	74.2	73.4
O_2	11.85	13.2
CO_2	4.22	6.6
H_2O	8.836	5.6
Ar	0.8932	1.3

Part of the power plant electricity production is used to compensate the power requirement in the allocation problem. This is assumed to cover any carbon dioxide emitted as part of the transportation, treatment and compression.

A6. Carbon Removal (Treatment)

Post combustion carbon removal using chemical based adsorption consists of two main elements, an absorber that separates carbon dioxide from the rest of the stream into a solvent and regeneration that removes carbon dioxide in concentrated form from the solvent. The process is illustrated below.

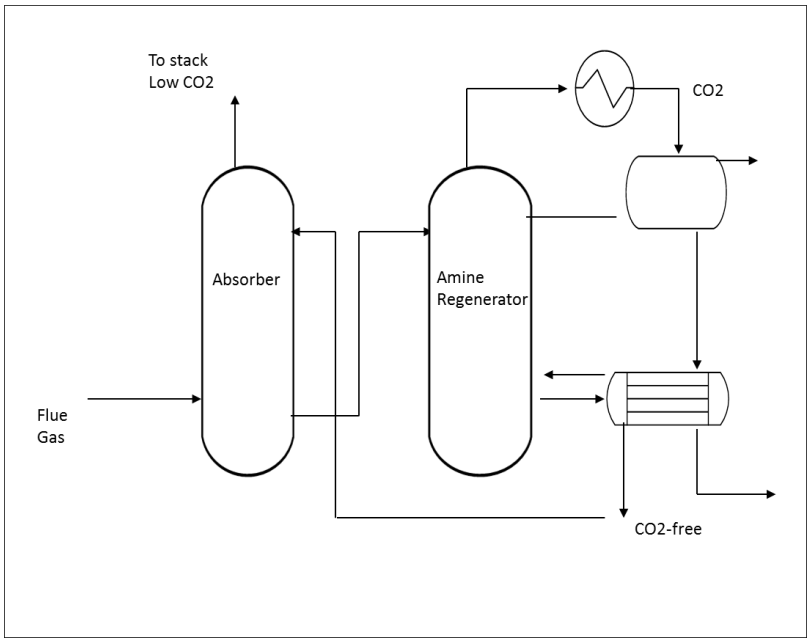

Figure 15: Amine Treatment Unit

Carbon dioxide recovery unit is able to recover certain amount of the inlet carbon dioxide due to process limitations. The remainder is unrecoverable and is often emitted to the atmosphere with any other contaminants that were present in the original stream. These include nitrogen oxides, sulfur oxides and water vapor along with the trace carbon dioxide [50] (Rao and Rubin, 2002[63], Rubin et al,2012[64]).

The recovery unit used in this work is assumed to be an amine unit, with an average recovery of 90%. However, the amine unit used also consumes power and heat. When power is already accounted for by the power station emissions, the heat is assumed to be generated locally.

Steam is used in the heat exchanger (re-boiler) at the regenerator and stripper units. It was reported that well designed Econamine FG plant have operated on less than 4.2 GJ/t CO_2, which is equivalent to 115 kg of 345 kPa saturated steam per metric cube solvent[50]. This required heat is associated with a carbon footprint. Using natural gas as the main fuel and assuming a heat exchanger efficiency of 85%, the emission rate was calculated as follows:

- Natural gas carbon dioxide footprint is 0.05 t CO2 emitted/GJ natural gas used
- Heat required per carbon dioxide is 4.2 GJ/t CO2 processed
- Re-boiler efficiency is 85%

Sample Calculation

$$Reboiler\ emission = \frac{0.05\ t\ CO_{2\ emitted}}{GJ\ Natural\ gas} * \frac{4.2\ GJ\ heat}{t\ CO_{2\ processed}} * \frac{1}{85\%}$$

$$Reboiler\ emission = \frac{0.25\ t\ CO_{2\ emitted}}{t\ CO_{2\ processed}}$$

This parameter was added to the overall recovery of the capture unit and giving an overall efficiency of

$$\varepsilon_T = (0.1 + 0.25)\frac{t\ CO_{2\ emitted}}{t\ CO_{2\ processed}}$$

$$\varepsilon_T = 0.35\ \frac{t\ CO_{2\ emitted}}{t\ CO_{2\ processed}}$$

APPENDIX B

Appendix B. Sink Processes and prices

B1. Methanol

In order for methanol reaction to take place, it requires 99.9% pure carbon dioxide. The capacity was chosen to be 1710 t/d CO_2 that needs to be supplied at 80MPa (Van Dal and Bouallou 2013, [65]). Hydrogen can be produced using water electrolysis unit powered by solar energy.

A lot of researches have been done in this area and most reported either a negative profit, defined the breakeven price of carbon dioxide subsided as taxes or aimed at maximizing intake in term of feasibility. Kim et al [66], economic evaluation of producing methanol from carbon dioxide using hydrogen powered by solar energy, assumes a price of 35 USD per ton CO_2. This was taken as the price methanol production would be able to provide to purchase carbon dioxide out of an amine treatment unit. Lower price was assumed in this work to take into account the competitive methanol market in the region. Since, the price of methanol today was at 482 USD per ton methanol [67], it safe to assume a value of 21 USD per ton CO_2 as an acceptable price to generate profit.

B2. Urea

Urea is assumed to be a sink and not a part of the plant sources emitting CO_2. Urea consumes 0.567 t NH3 per ton urea and 0.733 t CO_2 per ton urea. Using the excess

ammonia it can take extra 1126 CO_2 MTPD. Purity was assumed to be 99.9 wt% based on [58]. The exhaust is mainly dust and ammonia so it was assumed that it does not contain any carbon dioxide. Even though urea is sold at, 205 to 285 USD ton urea a price of 15 USD per ton CO_2 was assumed based on [68].

B3. EOR and Storage

The capital cost of enhanced oil recovery differs from one site to another depending on the location, type of well, CO_2 purchase price. Capital costs include compressors, separation equipment, well drilling and conversions and completions and H_2S removal. Operating costs include CO_2 purchase price that accounts for 55-75% of total costs, fuel costs and field operating [3]. The largest economic impacts are due oil prices and CO_2-EOR Operator [68].

Today, oil prices are at 100 USD bbl making carbon dioxide prices at 40-45 USD per ton CO_2 delivered [68]. A conservative price of 30 USD per ton CO_2 was used to account for the low end market demand. The requirement of pressure of 15 MPa [3] and purity of 94% which is the same as storage. Storage cost was assumed to be 8.6 USD/t CO_2 based on IPCC estimation [3]. Storage can take place in depleted oil filed, gas fields or in saline aquifers. Both the storage and EOR were assumed to have the same capacity.

B4. Algae

Algae can produce lipids, proteins and carbohydrates in large amounts in small time. They can be used in animal feed, as a food supplement and as bio-fertilizer. Even

though it has been under research since 1970's, it has not seen large commercial success yet. This is due to limitations in the processing section that consumes large amounts of energy. However, coupled with waste water treatment and with huge solar energy flux, algae can be used to sequestrate carbon dioxide and generate value added products. Since algae can take dilute carbon dioxide, it has been assumed to get it for free at no added cost. It is worth noting that algae production is highly dependent on land availability (Brennan and Owende, 2010 [59], Campbell et al, 2009 [70], Kumar et al, 2010[71]). Since land restrictions govern algae production, the carbon dioxide intake in the algae sink is limited by the available space within the industrial city. It is assumed that a free land plot of 1.44 km^2 exists. Based on Campbell et al [70] 1 ha pond is able to fix 185 t CO_2 per year and assuming an average productivity of 30 $g/m^2/d$, the algae production facility would be able to take 283 tons per year carbon dioxide

B5. Greenhouses

Carbon dioxide and water are the main chemicals needed for plants to sustain their lives along with soil nutrients and sunlight. Carbon dioxide and water are the building blocks for photosynthesis while light gives the needed energy to carry the reaction. While photosynthesis is mainly influenced by light intensity, it can improve with the increase of carbon dioxide concentration to a limit. This is all related each specific plant tolerance and chlorophyll level (Esmeijer, 1999[72]).

The use of carbon dioxide dosing in greenhouses can be supplied in two ways, with heat and without heat. Carbon dioxide with heat is generated through direct

combustion of fuel within the greenhouse facility to full the carbon dioxide dosing need and the required heating for the greenhouse. The second, without heat, means that while no heat is required in the greenhouse, carbon dioxide is still needed. This the case considered for a greenhouse based in Qatar. The high temperatures and the available sunlight throughout the year mean that there is no heat or sunlight shortage, quite the contrary, cooling might be required depending on the crops produced.

Carbon dioxide can be delivered then in two ways either dilute or pure without the need to consider heat transfer. Purification of flue gas, dilute stream of carbon dioxide, would be required to reduce NO_x that might harm some of crops. Moreover, it was reported that for the majority of the greenhouse crops, photosynthesis increases when levels of carbon dioxide was increase from 340 ppm to 1000 ppm. Liquid carbon dioxide popularity among growers have increased despite it being more expensive. The advantages of using pure carbon dioxide include ease of handling, less concern lacking crop contamination, no heat or moisture and the simple introduction to plants at any time or place. Pure liquid carbon dioxide has been estimated to cost around 5 USD/tons carbon dioxide [73]. The conditions thus were assumed to be the same as storage and EOR at 94 wt%. The size of the greenhouse and the subsequent amount of carbon dioxide intake is limited to land availability. Since, food production is subject to contamination, the greenhouse was located outside of the city boundaries. This was gave basis to assume a larger land plot would be available and thus larger carbon dioxide intake. The land area was assume to 12.2 km^2 and based on 0.5 kg CO_2/h/100 m^2, the capacity is calculated to be 1030 tons CO_2 per year.

B6. Efficiency factor for sink carbon release

Table 12: Sinks Efficiency Parameters

Sink	Capacity CO_2	Unit	Emitted CO_2	Unit	Source
Greenhouses	0.5	kg/h/100 m^2	0.5	t emitted/t supplementary	[73]
Algae	1.97	t CO_2/1 ha (pond)	0.42	%/yr of total sequestrated	[68]
EOR	8317	t/d, assumed	0.00076	(assumed)	[3]
Storage	8317	t/d, assumed	0	Calculated	[3]
Methanol	1710	t/d CO_2	0.098	t CO_2 emitted /t Fixed	[68]
Urea	1126	t/d CO_2	0.29		(Kojima et al, 2008[74])

APPENDIX C

Appendix C. Cost Correlations

C1. Pipeline

The pipeline expression implemented was based on a study carried by Parker [54]

Table 13: Pipeline Correlation Data

Diameter inches	Average cost, USD/mi
4	486492
6	467407
8	478076
10	503489
12	755993
16	855411
20	1055529
24	1210092
30	1469456
36	1768710
42	2301044

The data was fitted with the exception of the first two points as they showed non-linearity; as a result a slightly more expensive cost expression was reached to be Diameter Cost (USD/mi) = 50,193 Diameter (in) + 51,075

This linear cost expression as opposed to step-wise size based cost selection reduced the solution time and was able to approximate costs for installed pipelines. The data was obtained in the year 2000, therefore, using cost indices it was adjusted for inflation to the year 2009.

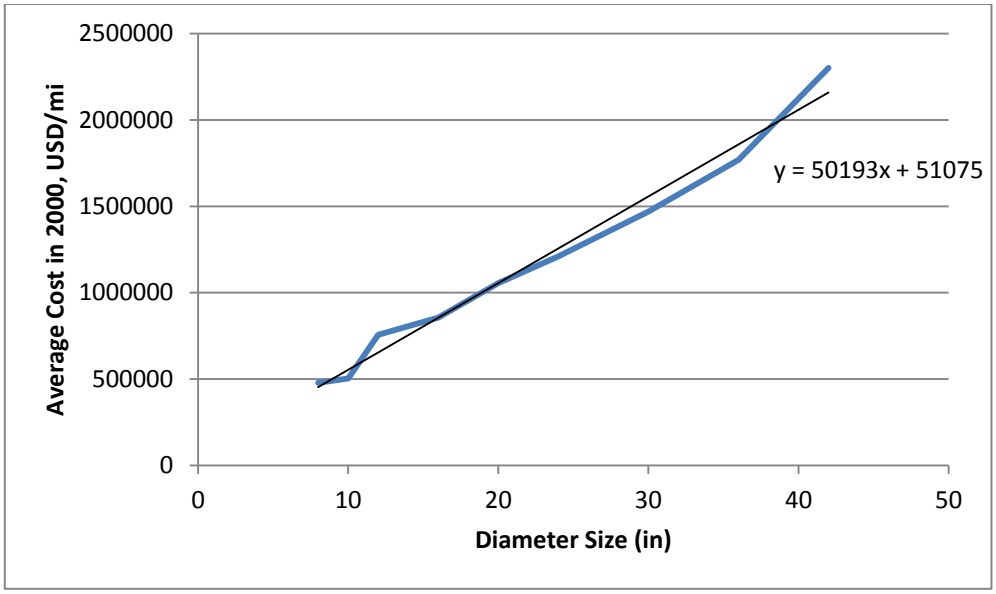

Figure 16: Linear Pipeline Cost Correlation

C2. Compression

Compression was simulated using Aspen Plus simulation software. The simulation was carried out in using a 4-stage centrifugal compressor with intercooling. The stream is compressed from its initial pressure to a cutoff pressure after which the gas transported enters the dense critical pressure phase. This change in phase requires a switch in equipment where a pump is used to reach the final pressure. This is done to avoid two-phase flow. The specific power for each connection is presented below

Table 14: Connections Specific Power in kWh/kg

Source/Sink	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia	0.010191	0.053179	0.10493	0.10493	0.10493	0.10493
Steel	0.011828	0.053484	0.10493	0.10493	0.10493	0.10493
Power plant	0.014807	0.054794	0.10493	0.10493	0.10493	0.10493
Refinery	0.013825	0.054598	0.10493	0.10493	0.10493	0.10493

The phase diagram below shows the area which the temperature and pressure of the critical phase highlighted in red. The pressure is estimated to be around 70 MPa and at temperature of 31 C.

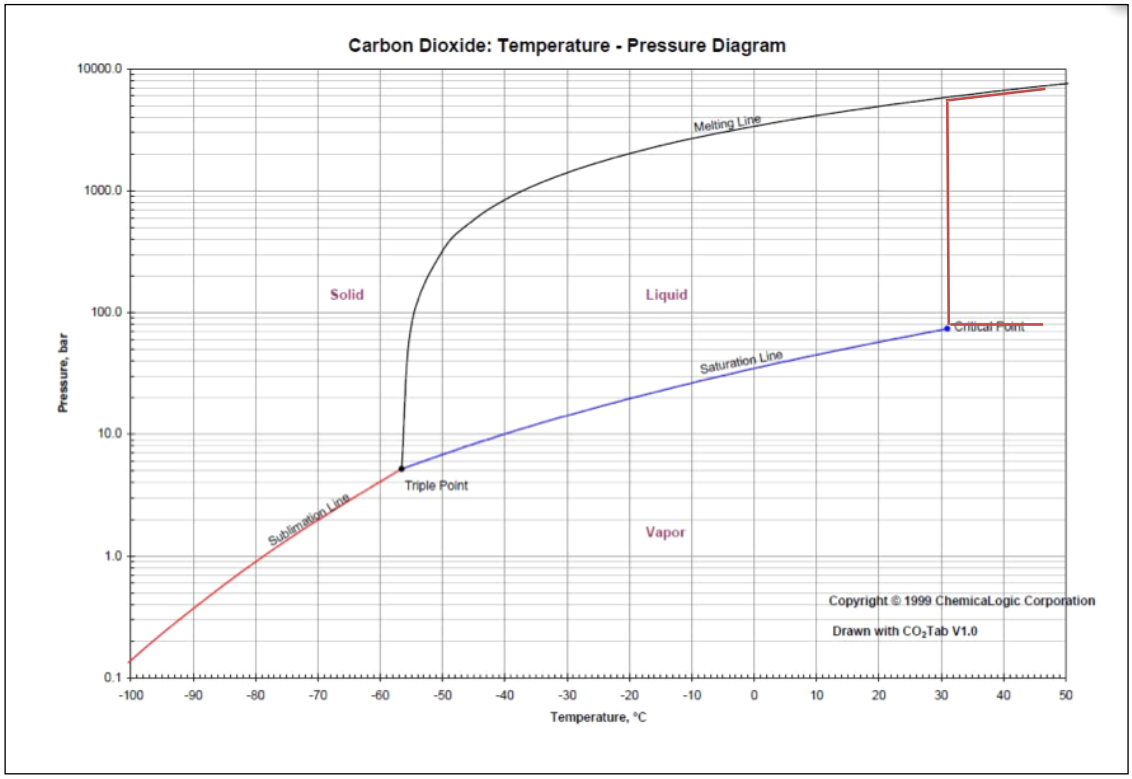

Figure 17: Carbon Dioxide Phase Diagram [75]

The compressor capital cost is estimated based on existing compressor cost with a known capacity raised to a specific exponent [51]. This value was brought to current cost using index values. The cost (USD/yr) with reference to the year 2004.

Compressor capital cost

$$= \text{ref. comp. cost}\left(\frac{\text{capacity of current comp.}}{\text{capacity of ref. comp. r}}\right)^{n}\frac{current\ index\ value}{ref.\ index\ value}$$

$$\text{Comp. capital cost (CC)} = 133000\left(\frac{\text{capacity of current comp.}}{224\ \text{kW}}\right)^{0.84}\frac{1194}{1000}$$

$$\text{CC}_{treated} = 158{,}902\left(\frac{\text{capacity of current comp.}}{224\ \text{kW}}\right)^{0.84}$$

Untreated flue gas is brought to the required pressure using a blower. This is due to the flue gas composition would be corrosive to the internals of a compressor.

The reference blower capacity was 472 m3/s for 67,000 USD and an n value of 0.6,

$$\text{CC}_{untreated} = 79{,}998\left(\frac{\text{capacity of current comp.}}{473\frac{\text{m3}}{\text{s}}}\right)^{0.6}$$

The capacity is the volumetric flow rate in m3/s calculated based on inlet conditions and discharge pressure.

$$Q\left(\frac{m^3}{s}\right) = F_{si,k} * \frac{RT}{PM} * \frac{1}{86.4}$$

C3. Capture/Removal Cost

The sources carbon removal cost were as follow;

Table 15: Sources Carbon Removal Costs

Plant	Sink	Composition wt%	Estimated C_{si}^{T}
Ammonia	Amine removal	100.0%	0
Steel	Iron production	44.0%	29
Power	Gas turbine	7.0%	43.15
Refinery	Boiler	27.0%	34.8

Most literature reported values for carbon dioxide removal include the cost of compression and dehydration. Since, compression is already accounted for in our model, that cost would be subtracted to obtain the removal cost only. Here it was conservatively assumed that the compression cost would be around 25% of the total cost obtained. A detailed cost breakdown of Hasan et al 2014 [41] supplementary material cites 80% of the total equipment cost for compression, making up around 46% of the overall cost. However, according to Rubin et al [64] compression would be responsible for around 30% as total energy penalty for a capture unit retrofitted to a power plant. They estimate that in newer plants, the cost of the capture and compression would be responsible for 80-90% penalty. Robeston [76] reported that compression and transportation cost would be responsible for 20% of the total cost of the capture. Moreover, Abu-Zahra et al [77] stated that the most expensive equipment of in an amine carbon removal unit is responsible of 55% of the total equipment purchased cost. Therefore, 25% was chosen as a safe in between estimate to be taken out of the capture cost reported in literature. The carbon removal cost parameter, C_{si}^{T} is presented in the table below.

Table 16: Literature Reported Carbon Removal Costs

Stream Name	Composition	Technology	Cost ,USD/t CO_2	Others	Source
Natural gas processing	-		30-70	Transport and Storage	Rubin et al, 2012[64]
Hydrogen production	-				
Ammonia production	-				
Ethylene oxide production	-				
Fischer-Tropsch Coal-to-liquid	-				
Biomass conversion	-		35-80	Transport and Storage	Rubin et al, 2012[64]
Biosynthetic gas	-				
ethanol production	-				
Biomass Hydrogen	-				
BTL	-				
Black liquor processing	-				
Refineries	-		45-120	Transport and Storage	Rubin et al, 2012[64]
Hydrogen from natural gas	-				
Methane or gasification residues	-				
FCC	-				
Process heaters	-				
Cement	-		55-150	Transport and Storage	Rubin et al, 2012[64]
Iron and steel	-		60-80	Transport and Storage	Rubin et al, 2012[64]
Coal plant Aus.	13% mol	MEA	88	compressed to 100 bars	Ho et al, 2013 [56]
Ammonia primary reformer stack			40 Capital 30 Oper.	incl. steam, cooling	Starit and Nagvekar, 2010 [57]
Cement plants	15-25%	Post Comb.	38		Hendrik and Grus, 2004[78]
Iron and Steel plants	15-20%	Post Comb.	40		Hendrik and Grus, 2004[78]
Ammonia plants flue gas	8%	Post Comb.	49		Hendrik and Grus, 2004[78]
ammonia plants pure CO_2	Pure	Post Comb.	4		Hendrik and Grus, 2004[78]

Table 16: Continued Literature Reported Carbon Removal Costs

Stream Name	Composition	Technology	Cost, USD/t CO_2	Others	Source
Corex	-	Membrane	18		Gielen, 2003[79]
Petrochemicals	20% mol	MEA	46	compressed to 110 bars	Farla et al, 1995[80]
Coal plant	12% vol		15		Metz et al , 2005[3]
Gas turbine	4% vol		75	net capture	Metz et al , 2005[3]
Coal plant	11.8% vol	MEA	42	none compression	Roberston, 2007[76]
Hydrogen or ammonia or gas processing	-	-	5-55	captured	Metz et al , 2005[3]
Blast Furnace (steel)	21% vol	MEA	70	compressed to 100 bars	Ho et al, 2011[81]
Corex	30% vol	MEA	51	compressed to 100 bars	Ho et al, 2011[81]
Refineries	3-18%	Post Comb.	39-57		Hendrik and Grus, 2004[78]
Hydrogen flue gas	8%	Post Comb.	49		Hendrik and Grus, 2004[78]
Hydrogen pure CO_2	Pure	Post Comb.	4.11		Hendrik and Grus, 2004[78]
Petrochemical plants	8-13%	Post Comb.	44-49		Hendrik and Grus, 2004[78] * 25% off, assuming it's compressed

APPENDIX D

Throughout this section, the treated sources streams are indicated by abbreviation T., while untreated sources are distinguished by the abbreviation as UT.

D1. CO_2 Exchange Streams where flows are given in terms of tons of CO_2 per day

Table 17: 635 t/d Carbon Dioxide Allocation

Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	977
Steel T.	0	0	0	0	0	0
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 18: 2050 t/d Carbon Dioxide Allocation

Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	976.9998723
Steel T.	0	0	0	0	0	2176.846085
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 19: 6000 t/d Carbon Dioxide Allocation

Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	977
Steel T.	0	0	0	0	0	3451
Power T.	0	0	0	0	0	3889
Refinery T.	0	0	0	1075.996377	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 20: 7542 t/d Carbon Dioxide Allocation

Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	951.9990622
Steel T.	0	0	0	0	0	3168
Power T.	0	0	1486.684621	1689.59948	0	4171.319803
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	282.9979678	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 21: 8186 t/d Carbon Dioxide Allocation

Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	900.5075348	0	0
Steel T.	0	0	0	0	0	3451
Power T.	0	0	3557.584422	0	0	4819.907641
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 22: 10,291 t/d Carbon Dioxide Allocation

Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	0
Steel T.	0	0	0	0	0	2705.857143
Power T.	0	0	7894.961241	0	0	1490.038759
Refinery T.	0	0	0	0	0	1092
Ammonia UT.	0	0	0	0	0	977
Steel UT.	0	0	415.6542264	0	0	329.4886307
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 23: 10,000 t/d Carbon Dioxide Allocation

Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	0
Steel T.	0	0	3450.849353	0	0	0
Power T.	0	0	2172.998945	0	0	7212.001055
Refinery T.	0	0	0	0	0	1092
Ammonia UT.	0	0	669.8359234	307.1640766	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

D2. Compression and Piping Costs, Capital and Operating where costs are given in US dollars per year

Table 24: Compression and Pumping Capital and Operating Costs for 635 t/d CO_2 (3%)

Capital Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	283597.2741
Steel T.	0	0	0	0	0	0
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	598697.0835
Steel T.	0	0	0	0	0	0
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Capital Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	43667.73871
Steel T.	0	0	0	0	0	0
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	27920.66221
Steel T.	0	0	0	0	0	0
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 25: Compression and Pumping Capital and Operating Costs for 2050 t/d CO_2 (10%)

Capital Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	283597.243
Steel T.	0	0	0	0	0	555860.6182
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	598697.0053
Steel T.	0	0	0	0	0	1333952.305
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Capital Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	43667.73438
Steel T.	0	0	0	0	0	84509.93868
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	27920.65856
Steel T.	0	0	0	0	0	62301.69973
Power T.	0	0	0	0	0	0
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 26: Compression and Pumping Capital and Operating Costs for 6000 t/d CO_2 (30%)

Capital Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	283597.2741
Steel T.	0	0	0	0	0	818586.1351
Power T.	0	0	0	0	0	905012.3191
Refinery T.	0	0	0	307547.1439	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	598697.0835
Steel T.	0	0	0	0	0	2114742.718
Power T.	0	0	0	0	0	2383145.3
Refinery T.	0	0	0	659361.2004	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Capital Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	43667.73871
Steel T.	0	0	0	0	0	127829.5164
Power T.	0	0	0	0	0	141940.8872
Refinery T.	0	0	0	13734.20846	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	27920.66221
Steel T.	0	0	0	0	0	98768.19828
Power T.	0	0	0	0	0	110647.1757
Refinery T.	0	0	0	2722.56251	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 27: Compression and Pumping Capital and Operating Costs for 7450 t/d CO_2 (36%)

Capital Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	277488.6959
Steel T.	0	0	0	0	0	761816.1302
Power T.	0	0	403510.1473	449292.3078	0	959887.5259
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	175.6307823	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	583376.7268
Steel T.	0	0	0	0	0	1941322.785
Power T.	0	0	911027.3985	1035371.833	0	2556148.414
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	104289.7722	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Capital Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	42818.99299
Steel T.	0	0	0	0	0	118207.884
Power T.	0	0	61266.86623	15636.66189	0	151482.7657
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	27206.18653
Steel T.	0	0	0	0	0	90668.69086
Power T.	0	0	42735.63947	4324.051234	0	118679.5462
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 28: Compression and Pumping Capital and Operating Costs for 8186 t/d CO_2 (40%)

Capital Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	264825.6175	0	0
Steel T.	0	0	0	0	0	818586.1351
Power T.	0	0	839771.2612	0	0	1083785.011
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	551823.1677	0	0
Steel T.	0	0	0	0	0	2114742.718
Power T.	0	0	2180056.776	0	0	2953597.387
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Capital Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	13051.90631	0	0
Steel T.	0	0	0	0	0	127829.5164
Power T.	0	0	131983.3395	0	0	173403.815
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	2148.199334	0	0
Steel T.	0	0	0	0	0	98768.19828
Power T.	0	0	102264.8941	0	0	137132.725
Refinery T.	0	0	0	0	0	0
Ammonia UT.	0	0	0	0	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 29: Compression and Pumping Capital and Operating Costs 10,000 t/d CO_2 (49%)

Capital Compression source/sink	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia	0	0	0	0	0	0
Steel	0	0	2933329	0	0	0
Power	0	0	1884128	0	0	5942239
Refinery	0	0	0	0	0	980553
Ammonia	0	0	12705483	3090866	0	0
Steel	0	0	0	0	0	0
Power	0	0	0	0	0	0
Refinery	0	0	0	0	0	0
Operating Compression Exchange	**Algae**	**Greenhouse**	**Storage**	**Methanol**	**Urea**	**EOR**
Ammonia T.	0	0	0	0	0	0
Steel T.	0	0	2114650.455	0	0	0
Power T.	0	0	1331594.845	0	0	4419451.379
Refinery T.	0	0	0	0	0	669168.081
Ammonia UT.	0	0	12709926.57	3089198.788	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Capital Pump Exchange	**Algae**	**Greenhouse**	**Storage**	**Methanol**	**Urea**	**EOR**
Ammonia T.	0	0	0	0	0	0
Steel T.	0	0	127799.674	0	0	0
Power T.	0	0	84702.92458	0	0	254252.0744
Refinery T.	0	0	0	0	0	47456.04716
Ammonia UT.	0	0	33235.16357	11370.45795	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Pump Exchange	**Algae**	**Greenhouse**	**Storage**	**Methanol**	**Urea**	**EOR**
Ammonia T.	0	0	0	0	0	0
Steel T.	0	0	98743.07694	0	0	0
Power T.	0	0	62464.15561	0	0	205190.9354
Refinery T.	0	0	0	0	0	31109.66708
Ammonia UT.	0	0	19138.50166	732.7530744	0	0
Steel UT.	0	0	0	0	0	0
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0

Table 30: Compression and Pumping Capital and Operating Cost for 10,291 t/d CO_2 (50%)

Capital Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	0
Steel T.	0	0	0	0	0	667308.5732
Power T.	0	0	1640456.163	0	0	404274.7095
Refinery T.	0	0	0	0	0	311384.962
Ammonia UT.	0	0	0	0	0	153.2141913
Steel UT.	0	0	221.1922804	0	0	192.4132192
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Compression Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	0
Steel T.	0	0	0	0	0	1658125.67
Power T.	0	0	4837963.547	0	0	913082.7631
Refinery T.	0	0	0	0	0	669168.081
Ammonia UT.	0	0	0	0	0	18540274.93
Steel UT.	0	0	34212358.74	0	0	27123030.62
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Capital Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	0
Steel T.	0	0	0	0	0	102495.6273
Power T.	0	0	280094.7988	0	0	60860.50824
Refinery T.	0	0	0	0	0	47456.04716
Ammonia UT.	0	0	0	0	0	43667.73871
Steel UT.	0	0	42610.7411	0	0	35959.51219
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0
Operating Pump Exchange	Algae	Greenhouse	Storage	Methanol	Urea	EOR
Ammonia T.	0	0	0	0	0	0
Steel T.	0	0	0	0	0	77442.08485
Power T.	0	0	226945.3874	0	0	42393.56658
Refinery T.	0	0	0	0	0	31109.66708
Ammonia UT.	0	0	0	0	0	27920.66221
Steel UT.	0	0	27030.87971	0	0	21431.86323
Power UT.	0	0	0	0	0	0
Refinery UT.	0	0	0	0	0	0